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OAQPS (Office of Air Quality Planning and Standards) Control Cost Manual (Fourth Edition)

(U.S.) Environmental Protection Agency, Research Triangle Park, NC

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OAQPS CONTROL COST MANUAL

Fourth Edition

EPA 450/3-90-006
January 1990

United States Environmental Protection Agency
Office of Air Quality Planning and Standards

Research Triangle Park, North Carolina 27711

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Abstract: The Manual compiles up-to-date capital costs, operating and maintenance expenses, and other costs for 'add-on' air pollution control systems and provides a comprehensive, concise, consistent, and easy-to-use procedure for estimating and (where appropriate) escalating these costs. ('Add-on' systems are those installed downstream of an air pollution source to control its emissions.) The Manual estimating procedure rests on the notion of the 'factored' or 'study' estimate, nominally accurate to within + or - 30%. This type of estimate is well suited to estimating control system costs intended for use in regulatory development. Study estimates are sufficiently accurate, yet do not require the detailed, site-specific data inputs needed to make definitive or other more accurate types of estimates.

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Chapter 1

INTRODUCTION

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Contents

1.1 Role of Cost in Setting of Regulations 1-2
1.2 Purpose of Manual 1-2
1.3 Organization of the Manual 1-3
1.4 Intended Users of the Manual 1-4
1.5 "Uniqueness" of the Manual 1-5
References 1-7

1.1 Role of Cost in Setting of Regulations

Cost has an important role in setting many state and federal air pollution control regulations. The extent of this role varies with the type of regulation. For some types of regulations, cost is explicitly used in determining their stringency. This use may involve a balancing of costs and environmental impacts, costs and dollar valuation of benefits, or environmental impacts and economic consequences of control costs.

For other types of regulations cost analysis is used to choose among alternative regulations with the same level of stringency. For these regulations, the environmental goal is determined by some set of criteria which do not include costs. However, cost-effectiveness analysis is employed to determine the minimum cost way of achieving the goal.

For some regulations, cost influences enforcement procedures or requirements for demonstration of progress towards compliance with an air quality standard. For example, the size of any monetary penalty assessed for noncompliance as part of an enforcement action needs to be set with awareness of the magnitude of the control costs being postponed by the noncomplying facility. For regulations without a fixed compliance schedule, demonstration of reasonable progress towards the goal is sometimes tied to the cost of attaining the goal on different schedules.

Cost is a vital input into two other types of analyses that also sometimes have a role in standard setting. Cost is needed for a benefit-cost analysis that addresses the economic efficiency of alternative regulations. Cost is also an input into any analysis of the economic impact of each regulatory alternative. An economic impact analysis deals with the consequences of the regulation for small businesses, employment, prices, and market and industry structure.

1.2 Purpose of Manual

The purpose of this *Manual* is two-fold: (1) to compile up-to-date capital costs, operating and maintenance expenses, and other costs for "add-on" air pollution control systems and (2) to provide a comprehensive, concise, con-

con't on pg
1-3

con't from pg 1-2

sistent, and easy-to-use procedure for estimating and (where appropriate) escalating these costs. ("Add-on" systems are those installed downstream of an air pollution source to control its emissions.)

+ or -

The *Manual* estimating procedure rests on the notion of the "factored" or "study" estimate, nominally accurate to within ± 30%. This type of estimate is well suited to estimating control system costs intended for use in regulatory development. Study estimates are sufficiently accurate, yet do not require the detailed, site-specific data inputs needed to make definitive or other more accurate types of estimates.

1.3 Organization of the Manual

This *Manual* is a revision of the 1987 edition of the *EAB Control Cost Manual*, [1] which, in turn, was a revision of the edition completed in 1978. This fourth edition of the *Manual* includes sizing and costing procedures and data for electrostatic precipitators and updated cost data for thermal and catalytic incinerators. This edition includes minor revisions to Chapters 2 ("Cost Estimating Methodology"), 5 ("Fabric Filters"), and 6 ("Electrostatic Precipitators") that serve to clarify the material therein. In addition, Chapter 3 ("Thermal and Catalytic Incinerators") has been rewritten to include device descriptions and sizing and costing procedures that are clearer, more concise, and more representative. Chapter 4 ("Carbon Adsorbers") has also been revised to add a comprehensive example problem and discussion on alternative methods for determining the carbon working capacity.

As with the third edition, this edition has been issued in self-contained chapters. Each chapter addresses a logically separate topic, which can be either of a general nature (*e.g.*, this introduction) or of a more specific, equipment-oriented nature (*e.g.*, fabric filters). The chapters which comprise this portion of the *Manual* are listed in Table 1.1, alongside the sections in the 1987 *Manual* they replace.

As in the third edition, each type of equipment, background topic, etc., is given its own chapter number, for ease of identification and to reinforce the intent that each chapter should stand alone. Also, each of the auxiliary equipment items (*e.g.*, ductwork) will be covered in a separate chapter.

Table 1.1: Contents of the *OAQPS Control Cost Manual* (Fourth Edition)

New Chapter		Old Section Replaced	
No.	Title	No.	Title
1	"Introduction"	1	"Introduction"
2	"Cost Estimating Methodology"	2	"Manual Estimating Methodology"
3	"Thermal and Catalytic Incinerators"	3	"Thermal and Catalytic Incinerators"
4	"Carbon Adsorbers"	4	"Carbon Adsorbers"
5	"Fabric Filters"	5	"Fabric Filters"
6	"Electrostatic Precipitators"	–	None

This has been done mainly to eliminate the confusion that arises when classifying auxiliaries like mechanical collectors which can either support a primary control device or be control devices in their own right.

Each of these stand-alone chapters contains a:

- *Process description*, where the types, uses, and operating modes of the equipment item and (if applicable) its auxiliaries are discussed;
- *Sizing procedure*, which enables one to use the parameters of the pollution source (*e.g.*, gas volumetric flow rate) to size the equipment item(s) in question;
- *Capital and annual costs* for the equipment and suggested factors to use in estimating these costs from equipment design and operational (*e.g.*, operating hours) parameters. These costs are presented in both graphical and equational forms wherever possible.

1.4 Intended Users of the Manual

As explained in Section 1.2, the *Manual* provides comprehensive procedures and data for sizing and costing control equipment. Some of these procedures are based on rigorous engineering principles, such as the material and engineering balances in Chapter 3. To fully appreciate, and correctly apply, these procedures the user should be able to understand them. Moreover, the user should be able to exercise "engineering judgement" on those occasions when the procedures may need to be modified or disregarded. Typically, only engineers and others with strong technical backgrounds possess this kind of knowledge. Hence, the *Manual* is oriented toward the *technical* not the non-technical user.

1.5 "Uniqueness" of the Manual

The *Manual* presents a different perspective on estimating air pollution control system costs than other cost-oriented reports, such as:

- *The Cost Digest: Cost Summaries of Selected Environmental Control Technologies* [2]
- *A Standard Procedure for Cost Analysis of Pollution Control Operations*[3]
- *Handbook: Control Technologies for Hazardous Air Pollutants* [4]

Although these reports (as well as many of the NSPS Background Information Documents) contain costs for add-on control systems, they do not duplicate the *Manual* for one or more of the following reasons: (1) their costs have been based either wholly or partly on data in the previous *Manuals*; (2) they apply to specific source categories only, whereas the *Manual* data may be applied generally; (3) their estimating procedures and costs are of less than study estimate quality; or (4) they are not intended for estimating costs used in regulatory development.

Reason (3) applies to the *Cost Digest*, for example, as this report, designed for use by non-technical personnel, contains procedures for making *order-of-magnitude* estimates (± 30% accuracy or worse). *A Standard Procedure*, conversely, was primarily intended for estimating costs for R&D cases (*e.g.*, demonstration projects), where some site-specific data are available. Further, although the latter report contains a thorough list of equipment installation factors, it contains few equipment *costs*. The report, *Handbook: Control Technologies*, used data and estimating procedures from the 1978 *Manual* to provide sound generalized procedures for estimating thermal and catalytic incinerator costs. The fourth edition of the *Manual* updates and expands this information.

Finally, since its inception, the *Manual* has been extensively used in Agency regulatory development efforts. Accordingly, the *Manual's* role in the speciality of air pollution control system cost estimating is both unique and secure.

References

[1] *EAB Control Cost Manual* (Third Edition), EPA, Office of Air Quality Planning and Standards, Economic Analysis Branch, February 1987 (EPA 450/5-87-001A [NTISPB 87-166583/AS]).

[2] DeWolf, Glenn, *et al.* (Radian, Inc.), *The Cost Digest: Cost Summaries of Selected Environmental Control Technologies*, EPA, ORD, Office of Environmental Engineering and Technology, October 1984 (EPA-600/884-010).

[3] Uhl, Vincent W., *A Standard Procedure for Cost Analysis of Pollution Control Operations*, Volumes I and II, EPA, ORD, Industrial Environmental Research Laboratory, June 1979 (EPA-600/8-79-018a).

[4] *Handbook: Control Technologies for Hazardous Air Pollutants*, EPA, ORD, Air and Energy Engineering Research Laboratory, September 1986 (EPA-625/6-86-014).

Chapter 2

COST ESTIMATING METHODOLOGY

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November 1989

Contents

2.1 Types of Cost Estimates 2-3
2.2 Cost Categories Defined 2-5
  2.2.1 Elements of Total Capital Investment 2-5
  2.2.2 Elements of Total Annual Cost 2-8
2.3 Engineering Economy Concepts 2-9
  2.3.1 Time Value of Money 2-9

2.3.2 Cash Flow 2-11

2.3.3 Annualization and Discounting Methods 2-14

2.4 Estimating Procedure 2-15

2.4.1 Facility Parameters and Regulatory Options 2-15

2.4.2 Control System Design 2-16

2.4.3 Sizing the Control System 2-18

2.4.4 Estimating Total Capital Investment 2-20

2.4.4.1 General Considerations 2-20

2.4.4.2 Retrofit Cost Considerations 2-22

2.4.5 Estimating Annual Costs 2-24

2.4.5.1 Raw Materials 2-24

2.4.5.2 Operating Labor 2-25

2.4.5.3 Maintenance 2-26

2.4.5.4 Utilities 2-26

2.4.5.5 Waste Treatment and Disposal 2-27

2.4.5.6 Replacement Parts 2-28

2.4.5.7 Overhead 2-28

2.4.5.8 Property Taxes, Insurance, and Administrative Charges 2-29

2.4.5.9 Capital Recovery 2-29

References 2-31

This chapter presents a methodology that will enable the user, having knowledge of the source being controlled, to produce study-level cost estimates for a control system to control that source. The methodology, which applies to each of the control systems included in this *Manual*, is general enough to be used with other "add-on" systems as well. Further, the methodology may also be applicable to estimating costs of fugitive emission controls and of other non-stack abatement methods.

Before presenting this methodology in detail, we should first discuss the various kinds of cost estimates and then define the cost categories and engineering economy concepts employed in making the estimates.

2.1 Types of Cost Estimates

As noted above, the costs and estimating methodology in this *Manual* are directed toward the "study" estimate, of ± 30% accuracy. According to Perry's *Chemical Engineer's Handbook*, a study estimate is "...used to estimate the economic feasibility of a project before expending significant funds for piloting, marketing, land surveys, and acquisition ... [However] it can be prepared at relatively low cost with minimum data."[1] Specifically, to make a study estimate, the following must be known:

- Location of the source within the plant;
- Rough sketch of the process flow sheet (*i.e.*, the relative locations of the equipment in the system);
- Preliminary sizes of, and material specifications for, the system equipment items;
- Approximate sizes and types of construction of any buildings required to house the control system;
- Rough estimates of utility requirements (*e.g.*, electricity);
- Preliminary flow sheet and specifications for ducting and piping;
- Approximate sizes of motors required.[1]

In addition, an estimate of the labor hours required for engineering and drafting is needed, as the accuracy of an estimate (study or otherwise) is highly dependent on the amount of engineering work expended on the project.

There are, however, four other types of estimates, three of which are more accurate than the study estimate. These are:[1]

- *Order-of-magnitude.* This estimate provides "a rule-of-thumb procedure applied only to repetitive types of plant installations for which there exists good cost history". Its error bounds are greater than ± 30%. (However, according to Perry's, "...no limits of accuracy can safely be applied to it.") The sole input required for making this level of estimate is the control system's capacity (often measured by the maximum volumetric flow rate of the gas passing through the system). So-called "six-tenths factor" estimates (not to be confused with *factored* estimates) are examples of this type.
- *Scope* or *Budget authorization* or *Preliminary.* This estimate, nominally of ± 20% accuracy, requires more detailed knowledge than the study estimate regarding the site, flow sheet, equipment, buildings, etc. In addition, rough specifications for the insulation and instrumentation are also needed.
- *Project control* or *Definitive.* These estimates, accurate to within ± 10%, require yet more information than the scope estimates, especially concerning the site, equipment, and electrical requirements.
- *Firm* or *Contractor's* or *Detailed.* This is the most accurate (± 5%) of the estimate types, requiring complete drawings, specifications, and site surveys. Further, "[t]ime seldom permits the preparation of such estimates prior to an approval to proceed with the project."[1]

For the purposes of regulatory development, study estimates have been found to be acceptable, as they represent a compromise between the less accurate order-of-magnitude and the more accurate estimate types. The former are too imprecise to be of much value, while the latter are not only very expensive to make, but require detailed site and process-specific knowledge that most *Manual* users will not have available to them.

2.2 Cost Categories Defined

The names given certain categories of costs and what they contain vary considerably throughout the literature. Certain words like "capital cost" can have vastly different meanings, which can often lead to confusion, even among cost estimators. To avoid this confusion and, at the same time, provide uniformity in the *Manual*, basic terms are defined in this chapter and will be used throughout. The terminology used is adapted from that of the American Association of Cost Engineers.[2] Although it has been developed for general use, it is readily adaptable to air pollution control system costing.

First, two general kinds of costs are estimated, ***total capital investment*** (TCI) and ***total annual cost*** (TAC). These are discussed below.

2.2.1 Elements of Total Capital Investment

The total capital investment includes all costs required to purchase equipment needed for the control system (termed *purchased equipment* costs), the costs of labor and materials for installing that equipment (termed *direct installation* costs), costs for site preparation and buildings, and certain other costs which are termed *indirect installation* costs. The TCI also includes costs for land, working capital, and off-site facilities.

Direct installation costs include costs for foundations and supports, erecting and handling the equipment, electrical work, piping, insulation, and painting. Indirect installation costs include such costs as engineering costs; construction and field expenses (*i.e.*, costs for construction supervisory personnel, office personnel, rental of temporary offices, etc.); contractor fees (for construction and engineering firms involved in the project); start-up and performance test costs (to get the control system running and to verify that it meets performance guarantees); and contingencies. Contingencies is a catch-all category that covers unforeseen costs that may arise, including (but certainly not limited to) "...possible redesign and modification of equipment, escalation increases in cost of equipment, increases in field labor costs, and delays encountered in start-up."[2]

These elements of total capital investment are displayed in Figure 2.1. Note that the sum of the purchased equipment cost, direct and indirect installation costs, site preparation, and buildings costs comprises the battery limits estimate. By definition, this is the total estimate "...for a specific job without regard to required supporting facilities which are assumed to already exist..."[2] at the plant. This would mainly apply to control systems installed in existing plants, though it could also apply to those systems installed in new plants when no special facilities for supporting the control system (*i.e.*, off-site facilities) would be required.

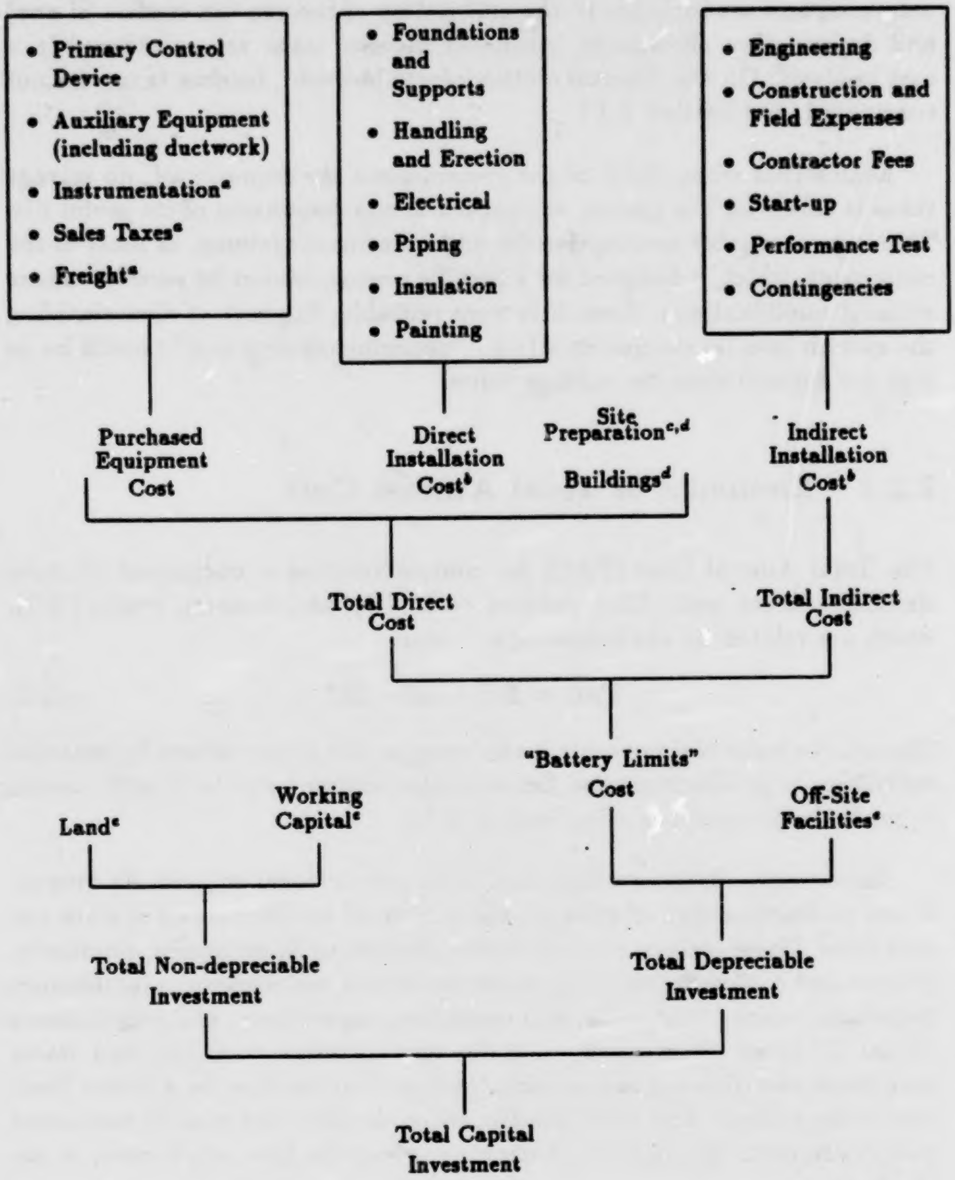
Where required, these off-site facilities would encompass units to produce steam, electricity, and treated water; laboratory buildings, railroad spurs, roads, and the like. It is unusual, however, for a pollution control system to have one of these units (*e.g.*, a power plant) dedicated to it. The system needs are rarely that great. However, it may be necessary—especially in the case of control systems installed in new or "grass roots" plants—for extra capacity to be built into the site generating plant to service the system. (A venturi scrubber, which often requires large amounts of electricity, is a good example of this.) It is customary for the utility costs to be charged to the project as operating costs at a rate which covers both the investment and operating costs for the utility.

As Figure 2.1 shows, there are two other costs which may be included in the total capital investment for a control system. These are *working capital* and *land*. The first of these, working capital, is a fund set aside to cover the initial costs of fuel, chemicals, and other materials, as well as labor and maintenance. It usually does not apply to control systems, for the quantities of utilities, materials, labor, etc., they require are usually small. (An exception might be an oil-fired thermal incinerator, where a small supply (*e.g.*, 30-day) of distillate fuel would have to be available during its initial period of operation.)

Land may also be required. But, since most add-on control systems take up very little space (a quarter-acre or less) this cost would be relatively small. (Certain control systems, such as those used for flue gas desulfurization, require larger quantities of land for the process equipment, chemicals storage, and waste disposal.)

Note also in Figure 2.1 that the working capital and land are nondepreciable expenses. In other words, these costs are "recovered" when the control system reaches the end of its useful life (generally in 10 to 20 years).

Figure 2.1: Elements of Total Capital Investment



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[a] Typically factored from the sum of the primary control device and auxiliary equipment costs.
[b] Typically factored from the purchased equipment cost.
[c] Usually required only at "grass roots" installations.
[d] Unlike the other direct and indirect costs, costs for these items usually are not factored from the purchased equipment cost. Rather, they are sized and costed separately.
[e] Normally not required with add-on control systems.

Conversely, the other capital costs are depreciable, in that they cannot be recovered and are included in the calculation of income tax credits (if any) and depreciation allowances, whenever income taxes are considered in a cost analysis. (In the *Manual* methodology, however, income taxes are not considered. See Section 2.3.)

Notice that when 100% of the system costs are depreciated, no salvage value is taken for the system equipment at the conclusion of its useful life. This is a reasonable assumption for add-on control systems, as most of the equipment, which is designed for a specific source, cannot be used elsewhere without modifications. Even if it were reusable, the cost of disassembling the system into its components (*i.e.*, "decommissioning cost") could be as high (or higher) than the salvage value.

2.2.2 Elements of Total Annual Cost

The Total Annual Cost (TAC) for control systems is comprised of three elements: *direct costs* (DC), *indirect costs* (IC), and *recovery credits* (RC), which are related by the following equation:

$$TAC = DC + IC - RC \tag{2.1}$$

Clearly, the basis of these costs is one year, as this period allows for seasonal variations in production (and emissions generation) and is directly usable in profitability analyses. (See Section 2.3.)

Direct costs are those which tend to be proportional or partially proportional to the quantity of exhaust gas processed by the control system per unit time. These include costs for raw materials, utilities (steam, electricity, process and cooling water, etc.), waste treatment and disposal, maintenance materials, replacement parts, and operating, supervisory, and maintenance labor. Of these direct costs, costs for raw materials, utilities, and waste treatment and disposal are *variable*, in that they tend to be a direct function of the exhaust flow rate. That is, when the flow rate is at its maximum rate, these costs are highest. Conversely, when the flow rate is zero, so are the costs.

Semivariable direct costs are only partly dependent upon the exhaust flow rate. These include all kinds of labor, maintenance materials, and replacement parts. Although these costs are a function of the gas flow rate, they are not linear functions. Even while the control system is not operating, some of the semivariable costs continue to be incurred.

Indirect, or "fixed", annual costs are those whose values are totally independent of the exhaust flow rate and, in fact, would be incurred even if the control system were shut down. They include such categories as overhead, administrative charges, property taxes, insurance, and capital recovery.

Finally, the direct and indirect annual costs are offset by recovery credits, taken for materials or energy recovered by the control system, which may be sold, recycled to the process, or reused elsewhere at the site. These credits, in turn, must be offset by the costs necessary for their processing, storage, transportation, and any other steps required to make the recovered materials or energy reusable or resalable. Great care and judgement must be exercised in assigning values to recovery credits, since materials recovered may be of small quantity or of doubtful purity, resulting in their having less value than virgin material.

The various annual costs and their interrelationships are displayed in Figure 2.2. A more thorough description of these costs and how they may be estimated is given in Section 2.4.

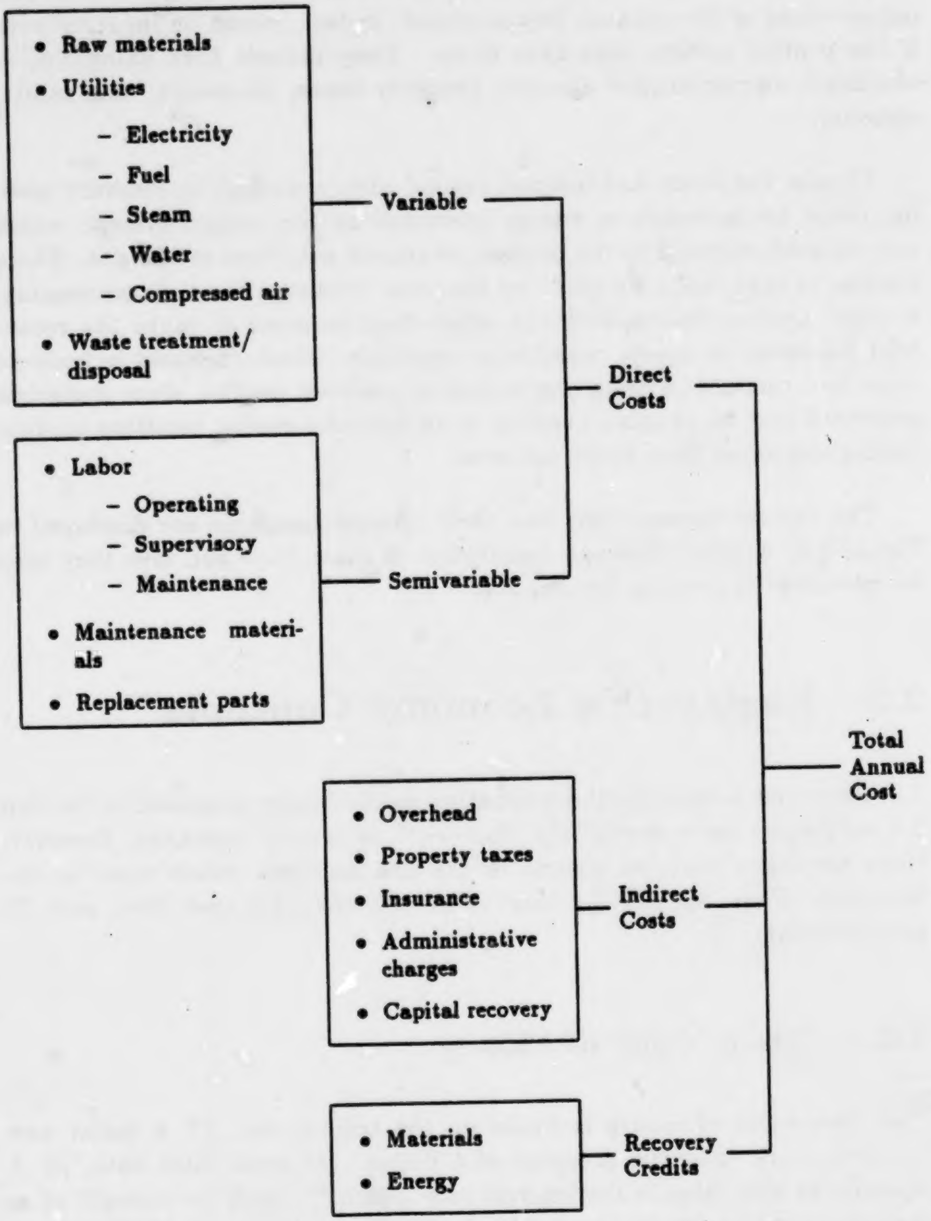
2.3 Engineering Economy Concepts

As mentioned previously, the estimating methodology presented in Section 2.4 rests upon the notion of the "factored" or "study" estimate. However, there are other concepts central to the cost analyses which must be understood. These are (1) the time value of money, (2) cash flow, and (3) annualization.

2.3.1 Time Value of Money

The *time value of money* is based on the truism that "...a dollar now is worth more than the prospect of a dollar...at some later date."[3] A measure of this value is the interest rate which "...may be thought of as the return obtainable by the productive investment of capital."[3]

Figure 2.2: Elements of Total Annual Cost



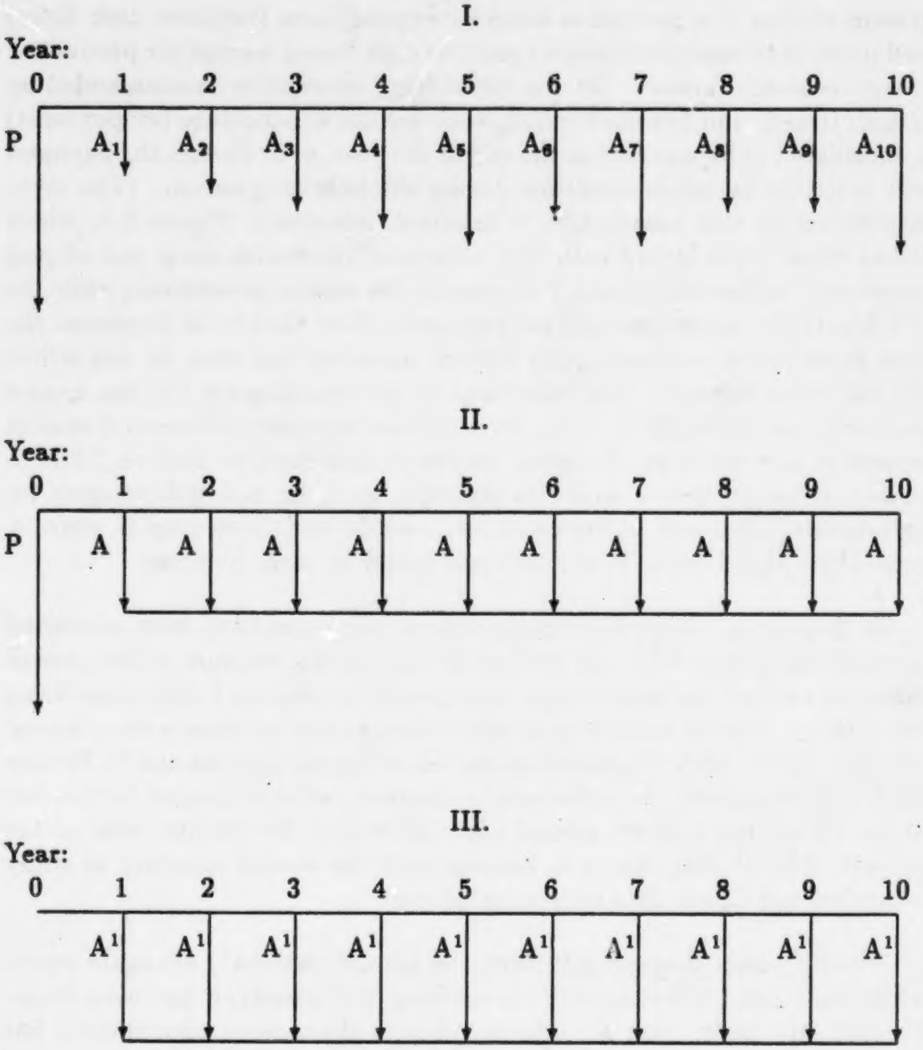
2.3.2 Cash Flow

During the lifetime of a project, various kinds of cash expenditures are made and various incomes are received. The amounts and timing of these expenditures and incomes constitute the *cash flows* for the project. In control system costing it is normal to consider expenditures (negative cash flows) and unusual to consider income (positive cash flows), except for product or energy recovery income. By the simplifying convention recommended by Grant, Ireson, and Leavenworth[3], each annual expenditure (or payment) is considered to be incurred at the end of the year, even though the payment will probably be made sometime during the year in question. (The error introduced by this assumption is minimal, however.) Figure 2.3, which shows three hypothetical cash flow diagrams, illustrates these end-of-year payments. In these diagrams, P represents the capital investment, while the A's denote the end-of-year annual payments. Note that in all diagrams, the cash flows are in *constant* (real) dollars, meaning that they do not reflect the effects of inflation. Also note that in the top diagram (I), the annual payments are different for each year. (These represent the control system annual costs (exclusive of capital recovery) described in Section 2.2.) In reality, these payments would be different, as labor and maintenance requirements, labor and utility costs, etc., would vary from year to year. A generally upward trend in annual costs would be seen, however.

In diagram II, these fluctuating annual payments have been converted to equal payments. This can be done by calculating the sum of the present values of each of the annual payments shown in diagram I and annualizing the total net present value to equivalent equal annual payments via a capital recovery factor. (See discussion in the following paragraphs and in Section 2.3.3.) Alternatively, it is adequate to choose a value of A equal to the sum of the direct and indirect annual costs estimated for the first year of the project. This assumption is in keeping with the overall accuracy of study estimates and allows for easier calculations.

Finally, notice diagram III. Here, the annual costs (A^1) are again equal, while the capital investment (P) is missing. Put simply, P has been incorporated into A^1, so that A^1 reflects not only the various annual costs but the investment as well. This was done by introducing another term, the *capital recovery factor* (CRF), defined as follows: "when multiplied by a present debt or investment , [the CRF] gives the uniform end-of-year payment necessary to repay the debt or investment in *n* years with interest

Figure 2.3: Hypothetical Cash Flow Diagrams[a]



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[a] **All Values Are Constant Year (Real) Dollars**

rate i."[3] The product of the CRF and the investment (P) is the capital recovery cost (CRC):

$$CRC = CRF \times P \tag{2.2}$$

where

$$CRF = \frac{i(1+i)^n}{(1+i)^n - 1} \tag{2.3}$$

Therefore, A^1 is the sum of A and the CRC, or:

$$A^1 = A + CRF \times P \tag{2.4}$$

In this context, n is the control system economic life, which, as stated above, typically varies from 10 to 20 years. The interest rate (i) used in this *Manual* is a pretax marginal rate of return on private investment of 10% (annual). This value, which could also be thought of as a "real private rate of return", is used in most of the OAQPS cost analyses and is in keeping with current OAQPS guidelines[4] and the Office of Management and Budget recommendation for use in regulatory analyses.[5]

It may be helpful to illustrate the difference between *real* and *nominal* interest rates. The mathematical relationship between them is straightforward:[4]

$$(1 + i_n) = (1 + i)(1 + r) \tag{2.5}$$

where

i_n, i = the annual nominal and real interest rates, respectively

r = the annual inflation rate

Clearly, the real rate does not consider inflation and is in keeping with the expression of annual costs in constant (*i.e.*, real) dollars.

The above procedure using the pre-tax marginal (or real) rate of return on private investment is the appropriate method for assessing the costs from the perspective of the entity having to install the pollution control equipment. For example, costs developed with the above procedure can appropriately be used for answering questions concerning the market response to regulation like price increases, quantity adjustments, and reduced profitability.

In an idealized economy with perfectly competitive and complete markets, this private cost and the social cost would be equal. However, in a more realistic economy in which allocation of resources is distributed by taxes, credit restrictions, and other market imperfections, the cost to society is different than the private costs for capital expenditures. The costs to society are the relevant costs for use in answering questions about economic efficiency. For example, benefit cost analysis and cost-effectiveness analyses should focus on cost to society, not just the cost to the entity facing additional pollution control costs.

EPA has adopted a new approach, a two-stage approach, to discounting for social costs. This new approach begins with the same capital recovery costs (CRC) described above using the same 10% pre-tax marginal rate of return on private investment. The second step of the two-stage approach involves "discounting" both direct and indirect annual costs and CRC back to an initial date ("year 0") using a consumption rate of interest of 3%. (See Section 2.3.3 for an explanation of the discounting concept.) This results in a relatively higher cost of capital from society's perspective than from the perspective of the entity facing additional control cost. A detailed explanation of this procedure and when it should be employed is beyond the scope of this document. A fuller explanation is given in draft EPA guidelines [6]. However it is mentioned here because the CRC and direct and indirect annual costs are inputs to the two-stage procedure and must be sufficiently itemized to allow use in the two-stage procedure.

2.3.3 Annualization and Discounting Methods

The above method of smoothing out the investment into equal end-of-year payments, is termed the *equivalent uniform annual cash flow* (EUAC) method.[3] In addition to its inherent simplicity, this method is very useful when comparing the costs of two or more alternative control systems (*i.e.*, those which are designed to control the same source to an equivalent degree). In fact, the EUAC's—or simply the total annual costs—of two competing systems may be compared even if both the systems have different economic lives, say 10 and 20 years. We recommend that the EUAC method be used for estimating control costs unless particular circumstances preclude its use.

Comparisons of systems with different economic lives cannot be made, however, using the other two annualization (*i.e.*, profitability analysis) methods—***present worth*** (PW) and ***internal rate of return*** (IRR). The present worth (or discounted cash flow) method involves the discounting of all cash flows occurring after year 0 (*i.e.*, the system startup date) back to year 0. These cash flows are discounted by multiplying each by a discount factor, $\frac{1}{(1+i)^m}$, where m is the number of years from year 0 to the year in which the cash flow is incurred. The sum of these discounted cash flows is then added to the capital investment to yield the present worth of the project. The alternative having the highest present worth would be selected (in control system costing this is usually a negative number). But when comparing the present worths of alternative systems, the system lifetimes must be equal for the comparison to be valid.[3]

The third annualization method, internal rate of return (IRR), is similar to the present worth method, in that it involves the discounting of a series of unequal cash flows. However, where with the PW method the interest rate, i, is set beforehand, in the IRR method the interest rate is solved for (usually via trial-and-error) after arbitrarily setting the PW to zero. When comparing alternative systems, the one with the highest "IRR" (interest rate) is selected.[3] But here again, the alternative systems compared must have equal economic lives.

2.4 Estimating Procedure

The estimating procedure used in the *Manual* consists of five steps: (1) obtaining the ***facility parameters*** and ***regulatory options*** for a given facility; (2) roughing out the control system design; (3) sizing the control system components; (4) estimating the costs of these individual components; and (5) estimating the costs (capital and annual) of the entire system.

2.4.1 Facility Parameters and Regulatory Options

Obtaining the facility parameters and regulatory options involves not only assembling the parameters of the air pollution source (*i.e.*, the quantity, temperature, and composition of the emission stream(s)), but also compiling data for the facility's operation. (Table 2.1 lists examples of these.)

Note that two kinds of facility parameters are identified—intensive and extensive. The former are simply those variables whose values are independent of quantity or dimensions—*i.e.*, the extent of the system. Conversely, extensive parameters encompass all size-dependent variables, such as the gas volumetric flow rate.

Like the facility parameters, the regulatory options are usually specified by others. These options are ways to achieve a predetermined emission limit. They range from no control to maximum control technically achievable. The option provided will depend, firstly, on whether the emission source is a stack (point source), a process leak (process fugitives source) or an unenclosed or partly enclosed area, such as a storage pile (area fugitives source). Stacks are normally controlled by "add-on" devices. As discussed above, this *Manual* will deal primarily with these add-on devices. (However, some of these devices can be used to control process fugitives in certain cases, such as a fabric filter used in conjunction with a building evacuation system.) Add-ons are normally used to meet a specified emission level, although in the case of particulate emissions, they may also be required to meet an opacity level.

2.4.2 Control System Design

Step 2—roughing out the control system design—first involves deciding what kinds of systems will be priced (a decision that will depend on the pollutants to be controlled, exhaust gas stream conditions, and other factors), and what auxiliary equipment will be needed. When specifying the auxiliary equipment, several questions need to be answered:

- What type of hood (if any) will be needed to capture the emissions at the source?
- Will a fan be needed to convey the exhaust through the system?
- Is a cyclone or another pre-cleaner needed to condition the exhaust before it enters the control device?
- Will the captured pollutants be disposed of or recycled? How will this be done?

Table 2.1: Facility Parameters and Regulatory Options

Facility Parameters

- Intensive
    - Facility status (new or existing, location)
    - Gas characteristics (temperature, pressure, moisture content)
    - Pollutant concentration(s) and/or particle size distribution
- Extensive
    - Facility capacity
    - Facility life
    - Exhaust gas flow rate
    - Pollutant emission rate(s)

Regulatory Options

- No control
- "Add-on" devices
    - Emission limits
    - Opacity limits
- Process modifications
    - Raw material changes
    - Fuel substitution
- Others
    - Coal desulfurization

- Can the on-site utility capacity (*e.g.*, electricity) accommodate the added requirements of the control system?

The kinds of auxiliary equipment selected will depend on the answers to these and other site-specific questions. However, regardless of the source being controlled, each system will likely contain, along with the control device itself, the following auxiliaries:

- *Hood*, or other means for capturing the exhaust;
- *Ductwork*, to convey the exhaust from the source to, through, and from the control system;
- *Fan system* (fan, motor, starter, inlet/outlet dampers, etc.), to move the exhaust through the system;
- *Stack*, for dispersing the cleaned gas into the atmosphere.

2.4.3 Sizing the Control System

Once the system components have been selected, they must be sized. Sizing is probably the most critical step, because the assumptions made in this step will more heavily influence the capital investment than any other. Before discussing how to size equipment, we need to define the term. For the purposes of this *Manual*, "sizing" is the calculation (or estimation) of certain critical design parameters for a control device against which the equipment cost of that device is most accurately correlated. For instance, the equipment cost of an electrostatic precipitator (ESP) is most often correlated with its collecting area. This, in turn, is a function of the exhaust volumetric flow rate, the overall collection efficiency, and the empirically-determined *migration velocity*, the ESP critical parameter. (Table 2.2 lists examples of these parameters. For a full description of the ESP sizing procedure, see Chapter 6.)

Also listed in Table 2.2 are general parameters which must also be specified before the purchased cost of the system equipment can be estimated. Note that, unlike the control device parameters, these may apply to any kind of control system. These parameters include materials of construction

Table 2.2: Examples of Typical Control Device Parameters [11]

General

- Material of construction: *e.g.*, carbon steel
- Insulated? Yes
- Economic life: 20 yr
- Redundancy[a]: none

Device-Specific

- Gas-to-cloth ratio ("critical parameter"): 3.0 to 1
- Pressure drop: 6.0 in w.g. (inches water gauge)
- Construction: standard (vs. custom)
- Duty: continuous (vs. intermittent)
- Filter type: shaker
- Bag material: polyester, 16-oz.

---

[a] Refers to whether there are any extra equipment items installed (*e.g.*, fans) to function in case the basic item becomes inoperative, so as to avoid shutting down the entire system.

(which may range from carbon steel to various stainless steels to fiberglass-reinforced polyester), presence or absence of insulation, and the economic or useful life of the system. As indicated in Section 2.3.2, this last parameter is required for estimating the annual capital recovery costs. The lifetime not only varies according to the type of the control system, but with the severity of the environment in which it is installed. (Representative values for the system life and the other control device parameters will be presented in those chapters of the *Manual* covering them.)

2.4.4 Estimating Total Capital Investment

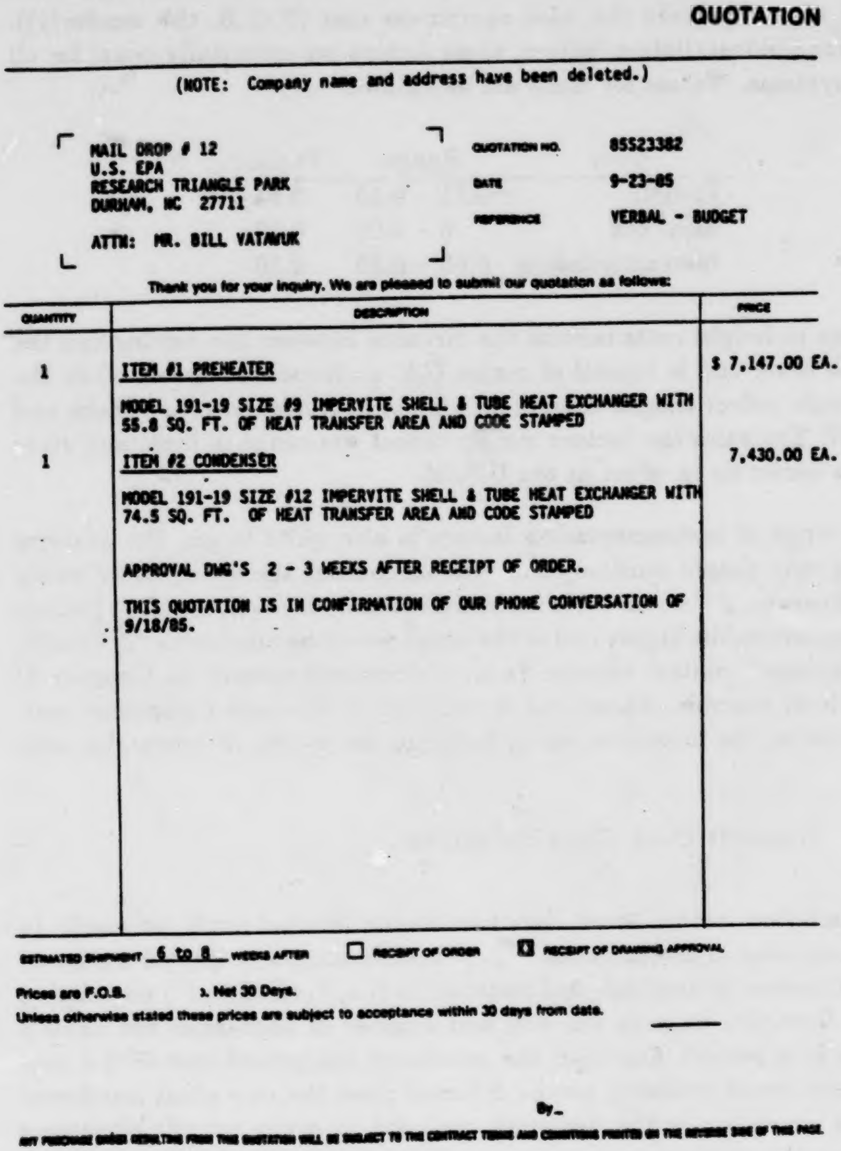
2.4.4.1 General Considerations

The fourth step is estimating the *purchased equipment cost* of the control system equipment. These costs are available from this *Manual* for the most commonly used add-on control devices and auxiliary equipment. Each type of equipment is covered in a separate chapter. (See Table of Contents.)

Most of these costs, in turn, have been based on data obtained from control equipment vendors. There are over one hundred of these firms, many of whom fabricate and erect a variety of control systems.[8] They have current price lists of their equipment, usually indexed by model designation. If the items for which costs are requested are fabricated, "off-the-shelf" equipment, then the vendor can provide a written quotation listing their costs, model designations, date of quotation, estimated shipment date, and other information. (See Figure 2.4 for a sample quotation.) Moreover, the quote is usually "F.O.B." (free-on-board) the vendor, meaning that no taxes, freight, or other charges are included. However, if the items are not off-the-shelf, they must be custom fabricated or, in the case of very large systems, constructed on-site. In such cases, the vendor can still give quotations—but will likely take much longer to do so and may even charge for this service, to recoup the labor and overhead expenses of his estimating department.

As discussed in Section 2.2 in this *Manual*, the total capital investment is factored from the purchased equipment cost, which in turn, is the sum of the base equipment cost (control device plus auxiliaries), freight, instrumentation, and sales tax. The values of these installation factors depend

Figure 2.4: Typical Vendor Quotation



QUOTATION

(NOTE: Company name and address have been deleted.)

MAIL DROP # 12
U.S. EPA
RESEARCH TRIANGLE PARK
DURHAM, NC 27711

ATTN: MR. BILL VATAVUK

QUOTATION NO. 85523382
DATE 9-23-85
REFERENCE VERBAL - BUDGET

Thank you for your inquiry. We are pleased to submit our quotation as follows:

QUANTITY	DESCRIPTION	PRICE
1	ITEM #1 PREHEATER MODEL 191-19 SIZE #9 IMPERVITE SHELL & TUBE HEAT EXCHANGER WITH 55.8 SQ. FT. OF HEAT TRANSFER AREA AND CODE STAMPED	$ 7,147.00 EA.
1	ITEM #2 CONDENSER MODEL 191-19 SIZE #12 IMPERVITE SHELL & TUBE HEAT EXCHANGER WITH 74.5 SQ. FT. OF HEAT TRANSFER AREA AND CODE STAMPED APPROVAL DWG'S 2 - 3 WEEKS AFTER RECEIPT OF ORDER. THIS QUOTATION IS IN CONFIRMATION OF OUR PHONE CONVERSATION OF 9/18/85.	7,430.00 EA.

ESTIMATED SHIPMENT 6 to 8 WEEKS AFTER ☐ RECEIPT OF ORDER ☒ RECEIPT OF DRAWING APPROVAL

Prices are F.O.B.). Net 30 Days.
Unless otherwise stated these prices are subject to acceptance within 30 days from date.

By_

ANY PURCHASE ORDER RESULTING FROM THIS QUOTATION WILL BE SUBJECT TO THE CONTRACT TERMS AND CONDITIONS PRINTED ON THE REVERSE SIDE OF THIS PAGE.

on the type of the control system installed and are, therefore, listed in the individual *Manual* chapters dedicated to them.

The costs of freight, instrumentation, and sales tax are calculated differently from the direct and indirect installation costs. These items are factored also, but from the base equipment cost (F.O.B. the vendor(s)). But unlike the installation factors, these factors are essentially equal for all control systems. Values for these are as follows:

Cost	Range	Typical
Freight	0.01 – 0.10	0.05
Sales tax	0 – 0.08	0.03
Instrumentation	0.05 – 0.30	0.10

The range in freight costs reflects the distance between the vendor and the site. The lower end is typical of major U.S. metropolitan areas, while the latter would reflect freight charges to remote locations such as Alaska and Hawaii.[7] The sales tax factors simply reflect the range of local and state tax rates currently in effect in the U.S.[9]

The range of instrumentation factors is also quite large. For systems requiring only simple continuous or manual control, the lower factor would apply. However, if the control is intermittent and/or requires safety backup instrumentation, the higher end of the range would be applicable.[7] Finally, some "package" control systems (*e.g.*, incinerators covered in Chapter 3) have built-in controls, whose cost is included in the base equipment cost. In those cases, the instrumentation factor to use would, of course, be zero.

2.4.4.2 Retrofit Cost Considerations

The installation factors listed elsewhere in the *Manual* apply primarily to systems installed in new facilities. These factors must be adjusted whenever a control system is sized for, and installed in (i.e, "retrofitted") an existing facility. However, because the size and number of auxiliaries are usually the same in a retrofit situation, the purchased equipment cost of the control system would probably not be different from the new plant purchased cost. An exception is the ductwork cost, for in many retrofit situations exceptionally long duct runs are required to tie the control system into the existing process.

Each retrofit installation is unique; therefore, no general factors can be developed. Nonetheless, some general information can be given concerning the kinds of system modifications one might expect in a retrofit:

1. *Auxiliaries.* Again, the most important component to consider is the ductwork cost. In addition, to requiring very long duct runs, some retrofits require extra tees, elbows, dampers, and other fittings.

2. *Handling and Erection.* Because of a "tight fit", special care may need to be taken when unloading, transporting, and placing the equipment. This cost could increase significantly if special means (*e.g.*, helicopters) are needed to get the equipment on roofs or to other inaccessible places.

3. *Piping, Insulation, and Painting.* Like ductwork, large amounts of piping may be needed to tie in the control device to sources of process and cooling water, steam, etc. Of course, the more piping and ductwork required, the more insulation and painting will be needed.

4. *Site Preparation.* Unlike the other categories, this cost may actually decrease, for most of this work would have been done when the original facility was built.

5. *Off-Site Facilities.* Conceivably, retrofit costs for this category could be the largest. For example, if the control system requires large amounts of electricity (*e.g.*, a venturi scrubber), the source's power plant may not be able to service it. In such cases, the source would have to purchase the additional power from a public utility, expand its power plant, or build another one. In any case, the cost of electricity supplied to that control system would likely be higher than if the system were installed in a new source where adequate provision for its electrical needs would have been made.

6. *Engineering.* Designing a control system to fit into an existing plant normally requires extra engineering, especially when the system is exceptionally large, heavy, or utility-consumptive. For the same reasons, extra supervision may be needed when the installation work is being done.

7. *Lost Production.* This cost is incurred whenever a retrofit control system cannot be tied into the process during normally scheduled

maintenance periods. Then, part or all of the process may have to be temporarily shut down. The revenue lost during this shutdown period is a bonafide retrofit expense.

8. *Contingency.* Due to the uncertain nature of many retrofit estimates, the contingency (*i.e.*, uncertainty) factor in the estimate should be increased.

From the above points, it is apparent that some or most of these installation costs would increase in a retrofit situation. However, there may be other cases where the retrofitted installation cost would be less than the cost of installing the system in a new plant. This could occur when one control device, say an ESP, is being replaced by a more efficient unit—a baghouse, for example. The ductwork, stack, and other auxiliaries for the ESP might be adequate for the new system, as perhaps would be the support facilities (power plant, etc.).

2.4.5 Estimating Annual Costs

Determining the total annual cost is the last step in the estimating procedure. As mentioned in Section 2.2 the TAC is comprised of three components—direct and indirect annual costs and recovery credits. Unlike the installation costs, which are factored from the purchased equipment cost, annual cost items are usually computed from known data on the system size and operating mode, as well as from the facility and control device parameters.

Following is a more detailed discussion of the items comprising the total annual cost. (Values/factors for these costs are also given in the chapters for the individual devices.)

2.4.5.1 Raw Materials

Raw materials are generally not required with control systems. Exceptions would be chemicals used in absorbers or venturi scrubbers as absorbents or to neutralize acidic exhaust gases (*e.g.*, hydrochloric acid). Chemicals may also be required to treat wastewater discharged by scrubbers or absorbers before releasing it to surface waters. But, these costs are only considered when a wastewater treatment system is exclusively dedicated to the control system. In most cases, a pro-rata waste treatment charge is applied. (See also discussion below on Waste Treatment and Disposal.)

Quantities of chemicals required are calculated via material balances, with an extra 10 to 20% added for miscellaneous losses. Costs for chemicals are available from the *Chemical Marketing Reporter* and similar publications.

2.4.5.2 Operating Labor

The amount of labor required for a system depends on its size, complexity, level of automation, and operating mode (*i.e.*, batch or continuous). The labor is usually figured on an hours-per-shift basis. As a rule, though, data showing explicit correlations between the labor requirement and capacity are hard to obtain. One correlation found in the literature is logarithmic:[10]

$$\frac{L_2}{L_1} = \left(\frac{V_2}{V_1}\right)^y \qquad (2.6)$$

where

L_1, L_2 = labor requirements for systems 1 and 2
V_1, V_2 = capacities of systems 1 and 2 (as measured by the gas flow rate, for instance)
y = 0.2 to 0.25 (typically)

The exponent in Equation 2.6 can vary considerably, however. Conversely, in many cases, the amount of operator labor required for a system will be approximately the same regardless of its size.

A certain amount must be added to operating labor to cover supervisory requirements. Fifteen per cent of the operating labor requirement is representative.[11]

To obtain the annual labor cost, multiply the operating and supervisory labor requirements by the respective wage rates (in $/hr) and the system operating factor (number of hours per year the system is in operation). The wage rates also vary widely, depending upon the source category, geographical location, etc. These data are tabulated and periodically updated by the U.S. Department of Labor, Bureau of Labor Statistics, in its *Monthly Labor Review* and in other publications. Finally, note that these are base labor rates, which do *not* include payroll and plant overhead. (See Overhead discussion below.)

2.4.5.3 Maintenance

Maintenance labor is calculated in the same way as operating labor and is influenced by the same variables. The maintenance labor rate, however, is normally higher than the operating labor rate, mainly because more skilled personnel are required. A 10% wage rate premium is typical.[11]

Further, there are expenses for maintenance materials—oil, other lubricants, duct tape, etc., and a host of small tools. Costs for these items can be figured individually, but since they are normally so small, they are usually factored from the maintenance labor. Reference [10] suggests a factor of 100% of the maintenance labor to cover the maintenance materials cost.

2.4.5.4 Utilities

This cost category covers many different items, ranging from electricity to compressed air. Of these, only electricity is common to all control devices, where fuel oil and natural gas are generally used only by incinerators; water and water treatment, by venturi scrubbers, quenchers, and spray chambers; steam, by carbon adsorbers; and compressed air, by pulse-jet fabric filters.

Techniques and factors for estimating utility costs for specific devices are presented in their respective sections. However, because nearly every system requires a fan to convey the exhaust gases to and through it, a general expression for computing the fan electricity cost (C_e) is given here:[7]

$$C_e = \frac{0.746\, Q\, \Delta P\, s\, \theta\, p_e}{6356\eta} \tag{2.7}$$

where

Q = gas flow rate (actual ft^3/min)

ΔP = pressure drop through system (inches of water, gauge) (Values for ΔP are given in the chapters covering the equipment items.)

s = specific gravity of gas relative to air (1.000, for all practical purposes)

θ = operating factor (hr/yr)

η = combined fan and motor efficiency (usually 0.60 to 0.70)

p_e = electricity cost ($/kwhr).

A similar expression can be developed for calculating pump motor electricity requirements.

2.4.5.5 Waste Treatment and Disposal

Though often overlooked, there can be a significant cost associated with treating and/or disposing of waste material captured by a control system that neither can be sold nor recycled to the process.

Liquid waste streams, such as the effluent from a venturi scrubber, are usually processed before being released to surface waters. The type and extent of this processing will, of course, depend on the characteristics of the effluent. For example, the waste can first be sent to one (or more) clarifiers, for coagulation and removal of suspended solids. The precipitate from the clarifier is then conveyed to a rotary filter, where most of the liquid is removed. The resulting filter cake is then disposed of, via landfilling, for example.

The annual cost of this treatment is relatively high—$1.00 to $2.00/thousand gallons treated or more.[12] The solid waste disposal costs (via landfilling, for example) typically would add another $20 to $30/ton disposed of.[13] This, however, would *not* include transportation to the disposal site. More information on these technologies and their costs is found in References [12] and [13].

2.4.5.6 Replacement Parts

This cost is computed separately from maintenance, because it is a large expenditure, incurred one or more times during the useful life of a control system. This category includes such items as carbon (for carbon adsorbers), bags (for fabric filters) and catalyst (for catalytic incinerators), along with the labor for their installation.

The annual cost of the replacement materials is a function of the initial parts cost, the parts replacement labor cost, the life of the parts, and the interest rate, as follows:

$$CRC_p = (C_p + C_{pl})\,CRF_p \tag{2.8}$$

where

CRC_p = capital recovery cost of replacement parts ($/yr)
C_p = initial cost of replacement parts, including taxes and freight ($)
C_{pl} = cost of parts-replacement labor ($)
CRF_p = capital recovery factor (defined in Section 2.3).

In the *Manual* methodology, replacement parts are treated the same as any other investment, in that they are also considered an expenditure that must be amortized over a certain period. Also, the useful life of the parts (typically 2 to 5 years) is generally less than the useful life of the rest of the control system.

Replacement-part labor will vary, depending upon the amount of the material, its workability, accessibility of the control device, and other factors.

2.4.5.7 Overhead

This cost is easy to calculate, but often difficult to comprehend. Much of the confusion surrounding overhead is due to the many different ways it is computed and to the several costs it includes, some of which may appear to be duplicative.

There are, generally, two categories of overhead, *payroll* and *plant*. Payroll overhead includes expenses directly associated with operating, supervisory, and maintenance labor, such as: workmen's compensation, Social Security and pension fund contributions, vacations, group insurance, and other fringe benefits. Some of these are fixed costs (*i.e.*, they must be paid regardless of how many hours per year an employee works). Payroll overhead is traditionally computed as a percentage of the total annual labor cost (operating, supervisory, and maintenance).

Conversely, plant (or "factory") overhead account for expenses not necessarily tied to the operation and maintenance of the control system, including: plant protection, control laboratories, employee amenities, plant lighting, parking areas, and landscaping. Some estimators compute plant overhead by taking a percentage of all labor plus maintenance materials [10], while others factor it from the total labor costs alone.[2]

For study estimates, it is sufficiently accurate to combine payroll and plant overhead into a single indirect cost. This is done in this *Manual*. Also, overhead is factored from the sum of all labor (operating, supervisory, and maintenance) plus maintenance materials, the approach recommended in reference [10]. The factors recommended therein range from 50 to 70% [10] An average value of 60% is used in this *Manual*.

2.4.5.8 Property Taxes, Insurance, and Administrative Charges

These three indirect operating costs are factored from the system total capital investment, and typically comprise 1, 1, and 2% of it, respectively. Taxes and insurance are self-explanatory. Administrative charges covers sales, research and development, accounting, and other home office expenses. (It should not be confused with plant overhead, however.) For simplicity, the three items are usually combined into a single, 4% factor. This value, incidentally, is standard in all OAQPS cost analyses.

2.4.5.9 Capital Recovery

As discussed in Section 2.3, the annualization method used in the *Manual* is the equivalent uniform annualized cost method. Recall that the cornerstone of this method is the capital recovery factor which, when multiplied by the total capital investment, yields the capital recovery cost. (See Equation 2.2.)

However, whenever there are parts in the control system that must be replaced before the end of its useful life, Equation 2.2 must be adjusted, to avoid double-counting.

That is:

$$CRC_s = CRF_s[TCI - (C_p + C_{pl})] \tag{2.9}$$

where

CRC_s = capital recovery cost for control system ($/yr)
TCI = total capital investment for entire system ($)
CRF_s = capital recovery factor for control system.

The term $(C_p + C_{pl})$ accounts for the cost of those parts (including taxes and freight) that would be replaced during the useful life of the control system and the labor for replacing them. Clearly, CRF_s and CRF_p will not be equal unless the control system and replacement part lives are equal.

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Chapter 3

THERMAL and CATALYTIC INCINERATORS

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Contents

3.1 Introduction 3-3

3.2 Process Description 3-4

3.2.1 Thermal Incinerators 3-7

3.2.1.1 Direct Flame Incinerators 3-9

3.2.1.2 Recuperative Incinerators 3-10

3.2.1.3 Regenerative Incinerators 3-11

3.2.2 Catalytic Incinerators 3-13

3.2.2.1 Fixed-Bed Catalytic Incinerators 3-15

3.2.2.2 Fluid-Bed Catalytic Incinerators 3-16

3.2.3 Other Considerations—(Packaged versus Field-Erected Units, Auxiliary Equipment) 3-17

3.2.3.1 Packaged vs. Field-Erected Units 3-17

3.2.3.2 Acid Gas Scrubbers 3-17

3.2.3.3 Heat Exchangers (Preheaters and other Waste Energy Recovery Units) 3-18

3.2.3.4 Other Auxiliary Equipment 3-18

3.2.4 Technology Comparison 3-19

3.3 General Treatment of Material and Energy Balances 3-21

3.4 Design Procedures 3-22

3.4.1 Steps Common to Thermal and Catalytic Units 3-23

3.4.2 Steps Specific to Thermal Units 3-28

3.4.3 Steps Specific to Catalytic Units 3-36

3.5 Cost Analysis 3-42

3.5.1 Estimating Total Capital Investment 3-42

3.5.1.1 Equipment Costs, EC 3-43

3.5.1.2 Installation Costs 3-51

3.5.2 Estimating Total Annual Cost 3-51

3.5.2.1 Direct Annual Costs 3-51

3.5.2.2 Indirect Annual Costs 3-57

3.5.3 Cost Comparison for Example Case 3-58

3.6 Acknowledgements 3-58

Appendix 3A – Properties of Selected Compounds 3-60

References 3-64

3.1 Introduction

Incineration, like carbon adsorption, is one of the best known methods of industrial gas waste disposal. Unlike carbon adsorption, however, incineration is an ultimate disposal method in that the objectionable combustible compounds in the waste gas are converted rather than collected. On the other hand, carbon adsorption allows recovery of organic compounds which may have more value as chemicals than just their heating value. A major advantage of incineration is that virtually any gaseous organic stream can be incinerated safely and cleanly, provided proper engineering design is used.

The particular application of both thermal and catalytic incineration to gaseous waste streams containing volatile organic compounds (VOC) is discussed here. The U.S. Environmental Protection Agency defines any organic compound to be a VOC unless it is specifically determined *not* to be a VOC. Indeed, a number of organics (*e.g.*, methane) are specified as not being VOCs. Although both VOC and non-VOC organic compounds are combustible and are therefore important in the design of the incinerator, this distinction is important since it is the control of VOCs that is regulated.

3.2 Process Description

Seldom is the waste stream to be combusted a single organic compound. Rather, it is common to have a complex mixture of organic compounds. This mixture is typically analyzed for carbon, hydrogen, oxygen, and other elements; and an empirical formula is developed which represents the mixture. Combustion of such a mixture of organic compounds containing carbon, hydrogen, and oxygen is described by the overall exothermic reaction:

$$C_xH_yO_z + \left[x + \frac{y}{4} - \frac{z}{2}\right] O_2 \Longrightarrow xCO_2 + \frac{y}{2}H_2O \qquad (3.1)$$

The complete combustion products CO_2 and H_2O are relatively innocuous, making incineration an attractive waste disposal method. When chlorinated or sulfur-containing compounds are present in the mixture, the products of complete combustion include the acid components HCl or SO_2, respectively, in addition to H_2O and CO_2. In general, these streams would require removal of the acid components by a scrubber unit, which could greatly affect the cost of the incineration system. (The sizing and costing of these scrubbers is covered in the "Wet Scrubbers" chapter of this *Manual.*)

The heart of an incinerator system is a combustion chamber in which the VOC-containing waste stream is burned. Since the inlet waste gas stream temperature is generally much lower than that required for combustion, energy must be supplied to the incinerator to raise the waste gas temperature. Seldom, however, is the energy released by the combustion of the total organics (VOCs and others) in the waste gas stream sufficient to raise its own temperature to the desired levels, so that auxiliary fuel (*e.g.*, natural gas) must be added.

The combustion of the waste gases may be accomplished in a thermal incinerator or in a catalytic incinerator. In the catalytic incinerator a catalyst is used to increase the rate of the combustion reaction, allowing the combustion to occur at lower temperatures. Because the catalytic process operates at a lower temperature than the thermal process, less auxiliary fuel may be required in the catalytic process to preheat the waste gas.

Auxiliary fuel requirements may also be decreased, and energy efficiency improved, by providing heat exchange between selected inlet streams and the effluent stream. The effluent stream containing the products of combustion, along with any inerts that may have been present in or added to the inlet streams, can be used to preheat the incoming waste stream, auxiliary air, or both via a "primary", or recuperative, heat exchanger. It is useful to define the fractional energy recovery by the preheater, or primary, heat exchanger as follows:

$$\begin{array}{l}\text{Fractional} \\ \text{Energy} \\ \text{Recovery}\end{array} = \frac{\text{Energy actually recovered from flue gas}}{\begin{array}{c}\text{Maximum energy recoverable if flue gas approaches} \\ \text{lowest temperature available to heat exchanger}\end{array}} \qquad (3.2)$$

The energy actually recovered, the numerator of Equation 3.2, is the increase in sensible heat of the gas, *i.e.*, waste gas or waste gas plus dilution air, being heated. The maximum energy recoverable would be the decrease in sensible heat of the flue gas, if it were cooled to the temperature of the incoming waste gas. While this maximum energy recovery would be attained only with a very large heat exchanger, the concept of fractional energy recovery is useful in expressing the extent of the improvement in energy efficiency using a "primary" heat exchanger.

Energy efficiency can be further improved by placing another ("secondary") exchanger downstream of the primary exchanger to recover additional energy from the effluent stream (*e.g.*, to generate low pressure process steam or hot water). However, secondary energy recovery is generally not used, unless there is a specific on-site use for the steam or hot water.

The majority of industrial gases that contain VOCs are dilute mixtures of combustible gases in air. In some applications, such as air oxidation processes, the waste gas stream is very deficient in oxygen. Depending on the oxygen content of the waste stream, auxiliary air may be required to combust the total organic content of the waste gas as well as any auxiliary fuel that has been used.

The concentration of combustible gas in the waste gas stream plays an integral role in the design and operation of an incinerator. From a cost standpoint, the amount of air in excess of the stoichiometric amounts should be minimized. For safety reasons, however, any mixture within the flammability limits, on either the fuel-rich or fuel-lean side of the stoichiometric mixture, presents an unacceptable fire hazard as a feed stream to the incinerator. The lower, or fuel-lean, explosive limit (LEL) of a given organic compound defines the minimum concentration of that compound in air that can produce more energy than is needed to raise its own temperature to the ignition point (*i.e.*, ignite). Similarly, the upper, or fuel-rich, explosive limit (UEL) represents the highest concentration of the organic in air that is ignitable. In the latter case, air is limiting the reaction. Both the LEL and the UEL are measured at ambient conditions. Empirically, it has been found that mixtures of hydrocarbons in air at their LEL have a heating value of approximately 50 Btu/scf.

Since the majority of industrial waste gases that contain VOCs are dilute mixtures of combustible gases in air, their heating value is low and their oxygen content exceeds that required to combust both the waste organics (VOCs and others) and the auxiliary fuel. If a waste gas above 50 percent LEL (about 25 Btu/scf) is encountered, it must be diluted to satisfy fire insurance regulations. Generally, the streams are brought to below 25 percent LEL, although concentrations from 25 percent to 50 percent are permitted provided the waste stream is continuously monitored by LEL monitors. Because air is the usual diluent gas, care must be taken with preheating the diluted stream so that it remains below about 1200°F. (See discussion below on preheating.) A table showing LEL, UEL, and heats of combustion for selected organic compounds is given in Appendix 3A.

The goal of any incineration system is to control the amount of VOCs released to the environment. Performance of a control device such as an incinerator can be described by a control efficiency, defined according to the following equation:

$$\text{Control Eff.,\%} = \left[\frac{\text{Inlet mass rate VOC} - \text{Outlet mass rate VOC}}{\text{Inlet mass rate VOC}}\right] \times 100 \quad (3.3)$$

It is important to note, however, that incomplete combustion of the inlet VOCs could result in the formation of other VOCs not originally present. For example, the incomplete oxidation of dichloroethane can yield vinyl chloride. Both of these compounds are VOCs. The definition given in Equation 3.3 would still be meaningful, however, as long as the newly formed VOC (*e.g.*, vinyl chloride) is detected. This situation necessitates the complete chemical analysis of the inlet and outlet gas streams to confirm compliance with State and Federal regulations.

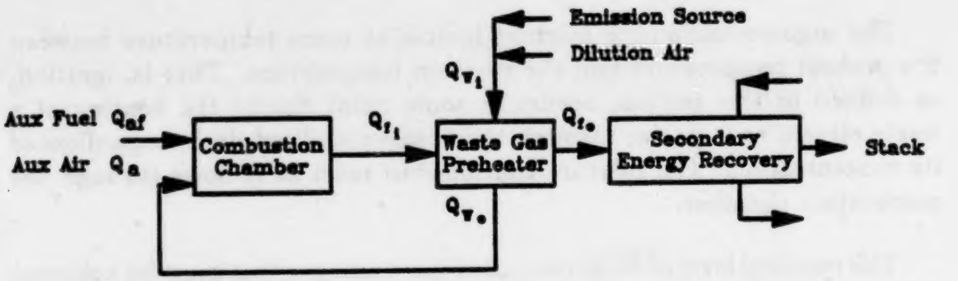

Figure 3.1: Thermal Incinerator – General Case

Performance of an incinerator can also be measured solely by the *outlet* VOC concentration, usually in ppmv.

There are a number of different incinerator designs. These designs can be broadly classified as thermal systems and catalytic systems. Thermal systems may be direct flame incinerators with no energy recovery, flame incinerators with a recuperative heat exchanger, or regenerative systems which operate in a cyclic mode to achieve high energy recovery. Catalytic systems include fixed-bed (packed-bed or monolith) systems and fluid-bed systems, both of which provide for energy recovery. The following sections discuss design aspects of these systems.

3.2.1 Thermal Incinerators

The heart of the thermal incinerator is a nozzle-stabilized flame maintained by a combination of auxiliary fuel, waste gas compounds and supplemental air added when necessary (see Figure 3.1). Upon passing through the flame, the waste gas is heated from its inlet temperature (*e.g.*, 100°F) to its ignition temperature. The ignition temperature varies for different compounds and is usually determined empirically. It is the temperature at

which the combustion reaction rate (and consequently the energy production rate) exceeds the rate of heat losses, thereby raising the temperature of the gases to some higher value. Thus, any organic/air mixture will ignite if its temperature is raised to a sufficiently high level.

The organic-containing mixture ignites at some temperature between the preheat temperature and the reaction temperature. That is, ignition, as defined in this section, occurs at some point during the heating of a waste stream as it passes through the nozzle-stabilized flame regardless of its concentration. The mixture continues to react as it flows through the combustion chamber.

The required level of VOC control of the waste gas that must be achieved within the time that it spends in the thermal combustion chamber dictates the reactor temperature. The shorter the residence time, the higher the reactor temperature must be. The nominal residence time of the reacting waste gas in the combustion chamber is defined as the combustion chamber volume divided by the volumetric flow rate of the gas. Most thermal units are designed to provide no more than 1 second of residence time to the waste gas with typical temperatures of 1,200 to 2,000°F. Once the unit is designed and built, the residence time is not easily changed, so that the required reaction temperature becomes a function of the particular gaseous species and the desired level of control. Table 3.1 illustrates the variability in (theoretical) reactor temperatures that is required to destroy 99.99 percent of the inlet mass of various noxious compounds with excess air for a 1-second reactor residence time [1].

These temperatures cannot be calculated a priori, although incinerator vendors can provide guidelines based on their extensive experience. In practice, most streams are mixtures of compounds, thereby further complicating the prediction of this temperature. Other studies [2,3,4], which are based on actual field test data, show that commercial incinerators should generally be run at 1600°F with a nominal residence time of 0.75 seconds to ensure 98% destruction of non-halogenated organics. In some States the reactor temperature and residence time of the unit are specified rather than attempting to measure actual levels of VOC control. The selected temperature must be maintained for the full, selected residence time for combustion to be complete.

These three studies also conclude that *mixing* is a critical factor in determining the destruction efficiency. Even though it cannot be measured,

Table 3.1: Theoretical Reactor Temperatures Required for 99.99 Percent Destruction by Thermal Incineration for a 1-Second Residence Time*

Compound	Temperature, °F
acrylonitrile	1,344
allyl chloride	1,276
benzene	1,350
chlorobenzene	1,407
1,2-dichloroethane	1,368
methyl chloride	1,596
toluene	1,341
vinyl chloride	1,369

*Reference [1]

mixing is a factor of equal or even greater importance than other parameters, such as temperature. The most feasible and efficient way to improve the mixing in an incinerator is to adjust it after start-up. The 98% control level discussed in the previous paragraph presumes such an adjustment.

Ultimately, once the unit is built, it is the responsibility of the user to operate and maintain the incinerator to insure compliance with applicable regulations.

3.2.1.1 Direct Flame Incinerators

Many configurations of thermal incinerators exist with the same goal—to raise the VOC-containing stream to the desired reaction temperature and hold it there for the given reaction time to achieve the required destruction efficiency. The simplest example of such a system is the direct flame incinerator. With reference to Figure 3.1, the direct flame incinerator is comprised only of the combustion chamber. The waste gas preheater and the secondary energy recovery heat exchanger are energy recovery devices and are not included as part of the direct flame incinerator.

3.2.1.2 Recuperative Incinerators

Recuperative incinerators have improved energy efficiency as a result of placing heat exchangers in the hot outlet gas streams. With reference to Figure 3.1, the recuperative incinerator is comprised of the combustion chamber, the waste gas preheater, and, if appropriate, the secondary energy recovery heat exchanger.

Primary Energy Recovery (Preheating Inlet Streams) Considerable fuel savings can be realized by using the exit (product) gas to preheat the incoming feed stream, combustion air, or both via a heat exchanger, as shown in Figure 3.1 in the so-called "recuperative" incinerator. These heat exchangers can recover up to 70% of the energy (enthalpy) in the product gas.

The two types of heat exchangers most commonly used are *plate-to-plate* and *shell-and-tube*. Plate-to-plate exchangers offer high efficiency energy recovery at lower cost than shell-and-tube designs. Also, because of their modular configuration, plate-to-plate units can be built to achieve a variety of efficiencies. But when gas temperatures exceed 1000°F, shell-and-tube exchangers usually have lower purchase costs than plate-to-plate designs. Moreover, shell-and-tube exchangers offer better long-term structural reliability than plate-to-plate units.[5] In any case, because most incinerators installed are packaged units, the design (and cost) of the recuperative heat exchangers have already been incorporated.

Most heat exchangers are not designed to withstand high temperatures, so that most of the energy needed to reach ignition is supplied by the combustion of fuel in the combustion chamber and only moderate preheat temperatures are sought in practice (<1200°F).

Secondary Energy Recovery (Additional Waste Energy Recovery) It should be noted, however, that at least some of the energy added by auxiliary fuel in the traditional thermal units (but not recovered in preheating the feed stream) can still be recovered. Additional heat exchangers can be added to provide process heat in the form of low pressure steam or hot water for on-site application. Obviously, an in-plant use for such low level energy is needed to realize these savings.

The need for this higher level of energy recovery will be dependent upon the plant site. The additional heat exchanger is often provided by the incineration unit vendor. The cost of this additional heat exchanger may be estimated via standard heat exchanger correlations and should be added to the costs estimated using the cost correlations in this chapter.

3.2.1.3 Regenerative Incinerators

A distinction in thermal incinerators can now be made on the basis of this limitation in the preheat temperature. The traditional approach to energy recovery in thermal units (shown schematically in Figure 3.1) still requires a significant amount of auxiliary fuel to be burned in the combustion chamber when the waste gas heating values are too low to sustain the desired reaction temperature at the moderate preheat temperature employed. Additional savings can, under these conditions, be realized in units with more complete transfer of exit stream energy. This is the concept behind the so-called excess-enthalpy or regenerable burner systems. These systems use direct contact heat exchangers constructed of a ceramic material that can tolerate the high temperatures needed to achieve ignition of the waste stream.

The operation of the regenerative system is illustrated in Figure 3.2. The inlet gas first passes through a hot ceramic bed thereby heating the stream (and cooling the bed) to its ignition temperature. If the desired temperature is not attainable, a small amount of auxiliary fuel is added in the combustion chamber. The hot gases then react (releasing energy) in the combustion chamber and while passing through another ceramic bed, thereby heating it to the combustion chamber outlet temperature. The process flows are then switched, now feeding the inlet stream to the hot bed. This cyclic process affords very high energy recovery (up to 95%).

The higher capital costs associated with these high-performance heat exchangers and combustion chambers may be offset by the increased auxiliary fuel savings to make such a system economical. The costs of these regenerative units will be given separately in the cost correlations presented in Section 3.5. Regenerative incinerators are not packaged units but are field-erected only. Accordingly, the costs given in Section 3.5 for regenerative units are for field-erected units.

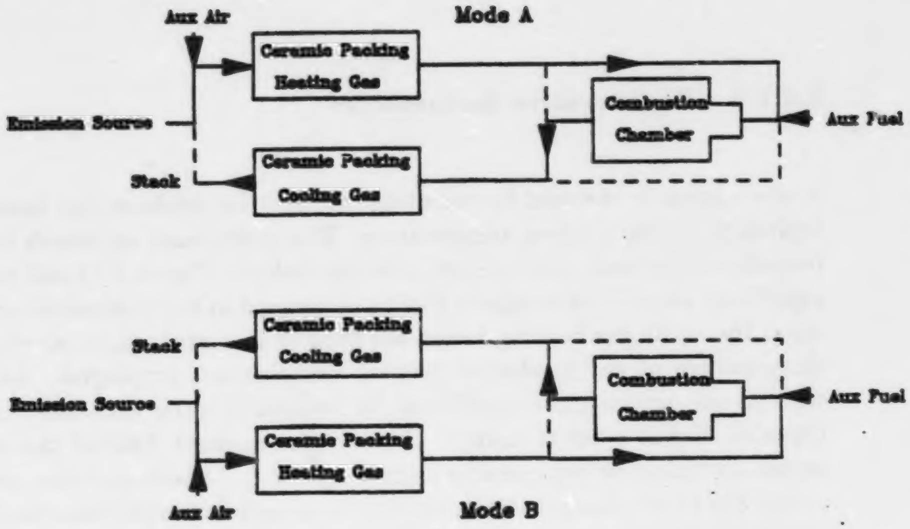

Figure 3.2: Regenerable-type Thermal Incinerator

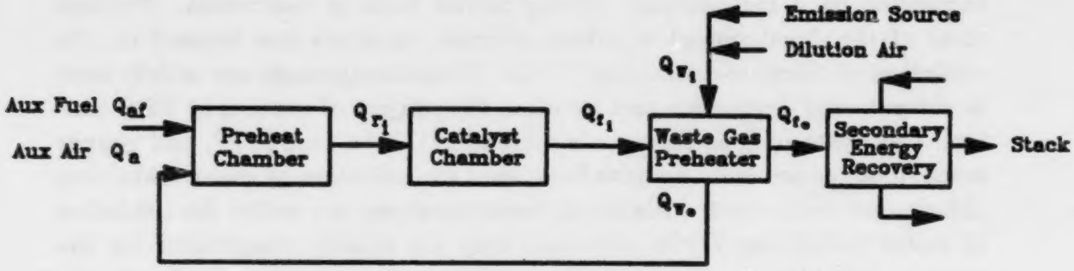

Figure 3.3: Catalytic Incinerator

3.2.2 Catalytic Incinerators

Catalytic incinerators employ a bed of active material (catalyst) that facilitates the overall combustion reaction given in Equation 3.1. The catalyst has the effect of increasing the reaction rate, enabling conversion at lower reaction temperatures than in thermal incinerator units. Nevertheless, the waste stream must be preheated to a temperature sufficiently high (usually from 300 to 900°F) to initiate the oxidation reactions. The waste stream is preheated either directly in a preheater combustion chamber or indirectly by heat exchange with the incinerator's effluent or other process heat or both (Figure 3.3). The preheated gas stream is then passed over the catalyst bed. The chemical reaction (combustion) between the oxygen in the gas stream and the gaseous pollutants takes place at the catalyst surface. Catalytic incineration can, in principle, be used to destroy essentially any oxidizable compound in an air stream. However, there are practical limits to the types of compounds that can be oxidized, due to the poisoning effect some species have on the catalyst. These limits are described below. In addition, most configurations require a low heating value of the inlet gas and a particulate content which is less than some small value.

Until recently, the use of catalytic oxidation for control of gaseous pollutants has generally been restricted to organic compounds containing only carbon, hydrogen and oxygen. Gases containing compounds with chlorine, sulfur, and other atoms that may deactivate the supported noble metal catalysts often used for VOC control were not suitably controlled by catalytic oxidation systems. Catalysts now exist, however, that are tolerant of such compounds. Most of these catalysts are single or mixed metal oxides, often supported by a mechanically strong carrier such as γ-alumina. Perhaps most of the development of poison-tolerant catalysts has focused on the oxidation of chlorine-containing VOCs. These compounds are widely used as solvents and degreasers and are often the subject of concern in VOC control. Catalysts such as chromia/alumina [6,7], cobalt oxide [8], and copper oxide/manganese oxide [9] have been used for oxidation of gases containing chlorinated compounds. Platinum-based catalysts are active for oxidation of sulfur-containing VOCs, although they are rapidly deactivated by the presence of chlorine. Compounds containing atoms such as lead, arsenic, and phosphorous should, in general, be considered poisons for most oxidation catalysts. Nevertheless, their concentration may be sufficiently low so that the rate of deactivation and therefore, the catalyst replacement costs, could be low enough to consider catalytic oxidation.

As was the case for thermal units, it is impossible to predict *a priori* the temperature and residence time (*i.e.*, inverse space velocity) needed to obtain a given level of conversion of a VOC mixture in a catalytic oxidation system. For example, Table 3.2 from Pope et al [8] shows the temperature needed for 80% conversion of a number of VOCs over two oxidation catalysts in a specific reactor design. This table shows that the temperature required for this level of conversion of different VOCs on a given catalyst and of the same VOC on different catalysts can vary significantly.

Particulate matter, including dissolved minerals in aerosols, can rapidly blind the pores of catalysts and deactivate them over time. Because essentially all the active surface of the catalyst is contained in relatively small pores, the particulate matter need not be large to blind the catalyst. No general guidelines exist as to particulate concentration and particulate size that can be tolerated by catalysts because the pore size and volume of catalysts vary greatly.

The volumetric gas flow rate and the concentration of combustibles in the gas flowing to the catalytic incinerator should be constant for optimal operation. Large fluctuations in the flow rate will cause the conversion

Table 3.2: Catalyst Temperatures Required for Oxidizing 80% of Inlet VOC to CO_2, °F for Two Catalysts

Compound	Temperature, °F	
	Co_3O_4	Pt - Honeycomb
acrolein	382	294
n-butanol	413	440
n-propylamine	460	489
toluene	476	373
n-butyric acid	517	451
1,1,1-trichloroethane	661	$\gg$661
dimethyl sulfide	—	512

of the VOCs to fluctuate also. Changes in the concentration or type of organics in the gas stream can also affect the overall conversion of the VOC contaminants. These changes in flow rate, organics concentration, and chemical composition are generally the result of upsets in the manufacturing process generating the waste gas stream. It may be uneconomical to change the process for the sake of making the operation of the catalytic incinerator feasible. In such cases, thermal incinerators (discussed earlier in this chapter) or carbon adsorption (discussed in Chapter 4 of this *Manual*) should be evaluated as alternative control technologies.

The method of contacting the VOC-containing stream with the catalyst serves to distinguish catalytic incineration systems. Both fixed-bed and fluid-bed systems are used.

3.2.2.1 Fixed-Bed Catalytic Incinerators

Fixed-bed catalytic incinerators may use a monolith catalyst or a packed-bed catalyst. Each of these is discussed below.

Monolith Catalyst Incinerators The most widespread method of contacting the VOC-containing stream with the catalyst is the catalyst monolith. In this scheme the catalyst is a porous solid block containing parallel, non-intersecting channels aligned in the direction of the gas flow. Monoliths offer the advantages of minimal attrition due to thermal expansion/contraction during startup/shutdown and low overall pressure drop.

Packed-Bed Catalytic Incinerators A second contacting scheme is a simple packed-bed in which catalyst particles are supported either in a tube or in shallow trays through which the gases pass. The first scheme is not in widespread use due to its inherently high pressure drop, compared to a monolith, and the breaking of catalyst particles due to thermal expansion when the confined catalyst bed is heated/cooled during startup/shutdown. However, the tray type arrangement, where the catalyst is pelletized is used by several industries (*e.g.*, heat-set web-offset printing). Pelletized catalyst is advantageous where large amounts of such contaminants as phosphorous or silicon compounds are present.[10]

3.2.2.2 Fluid-Bed Catalytic Incinerators

A third contacting pattern between the gas and catalyst is a fluid-bed. Fluid-beds have the advantage of very high mass transfer rates, although the overall pressure drop is somewhat higher than for a monolith. An additional advantage of fluid-beds is a high bed-side heat transfer as compared to a normal gas heat transfer coefficient. This higher heat transfer rate to heat transfer tubes immersed in the bed allows higher heat release rates per unit volume of gas processed and therefore may allow waste gases with higher heating values to be processed without exceeding maximum permissible temperatures in the catalyst bed. In these reactors the gas phase temperature rise from gas inlet to gas outlet is low, depending on the extent of heat transfer through imbedded heat transfer surfaces. The catalyst temperatures depend on the rate of reaction occurring at the catalyst surface and the rate of heat exchange between the catalyst and imbedded heat transfer surfaces.

As a general rule, fluid-bed systems are more tolerant of particulates in the gas stream than either fixed-bed or monolithic catalysts. This is due to the constant abrasion of the fluidized catalyst pellets, which helps remove these particulates from the exterior of the catalysts in a continuous manner.

A disadvantage of a fluid-bed is the gradual loss of catalyst by attrition. Attrition-resistant catalysts have been developed to overcome this disadvantage.[11]

3.2.3 Other Considerations—(Packaged versus Field-Erected Units, Auxiliary Equipment)

3.2.3.1 Packaged vs. Field-Erected Units

With the exception of regenerative incinerators, the equipment cost correlations included in this chapter are for packaged units only. They are *not* valid for field-erected units. For regenerative incinerators, the correlations are valid for field-erected units only. Packaged units are units that have been shop fabricated and contain all elements necessary for operation, except for connection to facilities at the site, *e.g.*, utilities. The elements include the combustion chamber, preheater, instrumentation, fan, and the necessary structural steel, piping, and electrical equipment. This equipment is assembled and mounted on a "skid" to facilitate installation on a foundation at the plant site. Tie-in to the local emission source is not part of the packaged unit. Units are usually sized to handle flow rates of <20,000 scfm, but can be built to accommodate flow rates up to 50,000 scfm. The cost correlations in this chapter are valid to 50,000 scfm for packaged units, except for fluid-bed units which are valid to 25,000 scfm.

Conversely, field-erected units may be built to any desired size. The combustion chamber, preheater, and other equipment items are designed and fabricated individually, and assembled at the site. However, both the equipment and installation costs of field-erected units are typically higher than those for equivalent-sized packaged units because the factors that improve efficiency of shop-fabrication, such as uniform working environment, availability of tools and equipment, and more efficient work scheduling, are generally not available in the field.

3.2.3.2 Acid Gas Scrubbers

The final outlet stream of any incineration system may contain certain pollutants that must be removed. The combustion of sulfur-containing compounds results in SO_2, while chlorinated compounds yield Cl_2 and HCl in the product stream. These acid gases must be removed from the gas stream if they are present at significant concentrations (regulations for limits on these gases vary from state to state). This removal can be effected in, for instance, a packed-bed vertical scrubber in which the flue gas is contacted with a caustic scrubbing liquid. For fluid-bed catalytic reactors, venturi scrubbers are often used because they provide for particulate removal as well as acid gas scrubbing. In most cases adding a scrubber significantly increases the cost of the incineration unit, sometimes by a factor of two. Costing of scrubbers is discussed in the "Wet Scrubbers" Chapter of this *Manual.*

If chlorinated VOCs are present in the waste gas, heat exchangers may require special materials of construction. This added expense is not included in the costing procedures outlined in this chapter.

3.2.3.3 Heat Exchangers (Preheaters and other Waste Energy Recovery Units)

For the thermal and catalytic units having some degree of energy recovery, the cost of the primary heat exchanger is included in the cost, and its design is usually done by the incineration unit vendor. The cost correlations presented in this chapter include units both with and without energy recovery. Secondary energy recovery, if desired, requires an additional heat exchanger, which is also often provided by the incineration unit vendor. Costing procedures for secondary energy recovery are not included in this chapter.

3.2.3.4 Other Auxiliary Equipment

Additional auxiliary equipment such as hoods, ductwork, precoolers, cyclones, fans, motors, and stacks are addressed separately in other chapters of this *Manual.*

3.2.4 Technology Comparison

Both the thermal and catalytic incineration systems are designed to provide VOC control through combustion at a level in compliance with applicable state and federal requirements. Given the wide range of options available, however, it is obvious that not every incinerator will fulfill these requirements at the same cost. This section presents a first step toward deciding how best to deal with VOC emission abatement using incinerators considering some qualitative factors pertinent to the types of incinerators described in this chapter. It is the intent of the remainder of Chapter 3 to provide a method by which the cost of VOC control for a particular application can be calculated.

A summary of the principal types of incinerators is presented in Table 3.3. From the earlier discussions, the following factors relating to the presence of contaminants should be considered by potential users [12]:

- The fouling of the catalyst in a catalytic system is a possibility. Poisons to the system include heavy metals, phosphorous, sulfur and most halogens, although catalysts have been developed that are chlorine resistant.

- The possibility of process upsets that could release any of the above poisons or cause fluctuations in the heating value to the incinerator would favor a thermal system.

- Except for the No.2 grade, fuel oil should not be considered as auxiliary fuel to a catalytic system due to the sulfur and vanadium it may contain.[10]

All of the above factors would serve to increase the operating expense of a catalytic unit through replacement costs of the catalyst. An additional factor relates to relative energy efficiency of the various types of incinerators:

- Thermal units generally require more auxiliary fuel than catalytic units and operate at temperatures that are roughly 1000°F higher. This difference in fuel requirement increases as the heating value of the waste stream decreases.

Table 3.3: Principal VOC Incinerator Technologies

Thermal Systems

- **Direct Flame Incinerator**
- **Recuperative Incinerator (Direct Flame with Recuperative Heat Exchanger)**
- **Regenerative Incinerator Operating in a Cyclic Mode**

Catalytic Systems

- **Fixed-Bed**
  - **Monolith**
  - **Packed-Bed**
- **Fluid-Bed**

In general, a trade-off exists between the higher capital costs of catalytic incinerators and the higher operating costs of thermal incinerators. This difference will be illustrated by a design example presented in Section 3.4 which treats both technologies.

3.3 General Treatment of Material and Energy Balances

In the sizing and costing of the incinerator and the calculation of the auxiliary fuel requirements, it is necessary to make material and energy balances around the entire incinerator unit and around certain parts of the unit, such as the combustion chamber or the preheater. This section presents a general approach to making these balances.

These balances are based on the law of conservation of mass and energy. They can be stated in general equation form as

$$In - Out + Generation = Accumulation \tag{3.4}$$

Because the incineration process is a steady-state process, the accumulation term is zero and the equation becomes

$$In - Out + Generation = 0$$

For *mass* balances it is useful to restrict the balances to be made on the mass of each atomic species so that for mass balances the generation term becomes zero. However, because the combustion reaction liberates energy, the *energy* balances around equipment where combustion takes place would include a generation term. Thus, the simplified equations are

$$In - Out = 0 \text{ , for steady-state mass balances} \tag{3.5}$$

$$In - Out + Generation = 0 \text{ , for steady-state energy balances} \tag{3.6}$$

For the incineration process the two terms *In* and *Out* are generally mass terms (for a mass balance) of the form

$$\rho Q,$$

where

ρ = density (mass per unit volume)
Q = volumetric flow rate (volume per unit time)

or sensible heat terms (for an energy balance) of the form,

$$\rho Q C_p (T - T_{ref})$$

where

C_p = heat capacity
T = temperature.

The reference temperature, T_{ref}, is often taken to be zero or the temperature of a convenient stream, *e.g.*, the inlet gas stream, in whatever units T is in, so the T_{ref} term may not appear in the equations. When the reference temperature is taken as zero, the sensible heat terms become

$$\rho Q C_p T.$$

Energy losses, H_L, are also part of the *Out* term and, for the incinerator process, are taken here to be 10% of the total energy *input* to the incinerator.

For the incineration process, the generation term for energy balances accounts for the energy released through the combustion reactions. This term is generally of the form

$$\rho Q(-\Delta h_c)$$

where

$(-\Delta h_c)$ = heat of combustion.

3.4 Design Procedures

The following procedure is designed to provide parameters for use in developing a study cost estimate (accuracy ± 30%). The principal parameters

of interest are

- **flue gas flow rate, upon which all the equipment cost correlations are based.**
- **auxiliary fuel requirement, which is important in estimating annual operating costs.**

For applications which involve control of waste gas streams that are *dilute mixtures of VOCs in air* (>20% oxygen in the waste gas stream), the flue gas flow rate is greater than the inlet waste gas flow rate by the amount of auxiliary fuel and the increase in the moles of gas as a result of the combustion reaction. Because these two factors usually cause only small increases in flow rate, a number of simplifying assumptions can be made in the design calculations. For applications where diluent air must be used to adjust the combustible concentration in the waste gas to 25% LEL and where auxiliary fuel and auxiliary combustion air are needed, more complete mass and energy balances must be made.

The design procedure illustrated below is for waste gas streams that are dilute mixtures of VOCs in air (>20% oxygen in the waste gas stream). In this discussion the design procedure will be illustrated by a sample problem that will be solved step-by-step.

3.4.1 Steps Common to Thermal and Catalytic Units

Step 1 - Establish design specifications The first step in the design procedure is to determine the specifications of the incinerator and the waste gas to be processed. The following parameters of the waste gas stream at the emission source must be available:

- Volumetric flow rate, scfm—Standard conditions are normally 77°F and 1 atm. pressure.
- Temperature
- Oxygen content
- Chemical composition of the combustibles

- Inerts content
- Heating value—In some cases the heating value may act as a surrogate for the chemical composition of the combustibles. This is particularly true for dilute mixtures of combustibles in air.
- Particulate content—The particulate content is important if catalytic incinerators are to be costed. An upstream filter may suffice if particulate content is too high. Fluid-bed catalytic incinerators can tolerate higher particulate contents than fixed-bed catalytic incinerators.

The following parameters must be specified for the incinerator:

- Desired control efficiency – This efficiency should be based on requirements dictated by relevant state and federal regulations.
- Combustion chamber outlet temperature – This temperature may also be based on requirements of a regulation or on recommendations developed during regulatory development.
- Desired percent energy recovery – The desired percent energy recovery should be the result of a process optimization in which costs of incinerators with several different levels of energy recovery are estimated and the minimum cost design selected. The tradeoff is between the capital cost of the energy recovery equipment and the operating (fuel) costs.

Specifications for the sample problem are given in Table 3.4.

Step 2 - Verify that the oxygen content of the waste gas exceeds 20% There must be sufficient oxygen in the waste gas to support the combustion of the waste organics (including VOCs) and the auxiliary fuel, if auxiliary fuel is needed. It may be necessary to add auxiliary air if the oxygen content is less than about 20%. This example is based on streams that contain >20% oxygen, as shown below:

$$\text{Air Content, Vol. \%} = 100.0 - \frac{1000}{10^6} \times 100 - \frac{1000}{10^6} \times 100 \tag{3.7}$$
$$= 99.8\%$$
$$\text{Oxygen Content,\%} = \text{Air Content} \times 0.209 \tag{3.8}$$
$$= 20.86\%$$

Table 3.4: Specification of Sample Problem

Variable	Value
Preheater Inlet Waste Gas Vol Flow Rate, Q_{w_i}, scfm	20,000
Preheater Inlet Waste Gas Temp., T_{w_i}, °F	100
Composition	
Benzene Content, ppmv	1000
Methyl Chloride Content, ppmv	1000
Air Content	Balance
Particulate Content	Negligible
Moisture Content	Negligible
Desired Control Efficiency,%	98
Desired Percent Energy Recovery, HR,%	70

Step 3 - Calculate the LEL and the Percent of the LEL of the gas mixture Note: If the waste stream contains a significant amount of inerts in addition to the nitrogen associated with the oxygen in air, the calculation of LEL (and UEL) loses meaning since the LEL (and UEL) is measured in mixtures of organic with air only. A complete chemical analysis is necessary to complete the design procedure in such a case.

The example chosen here is typical, in that there is more than one VOC component in the gas stream. An approximate method to calculate the LEL of a mixture of compounds, LEL_{mix}, is given by Grelecki [13] as

$$\text{LEL}_{mix} = \left[\sum_{j=1}^{n} \frac{x_j}{(\sum_{i=1}^{n} x_i) \times \text{LEL}_j}\right]^{-1} \tag{3.9}$$

where

x_i = volume fraction of combustible component i

LEL_j = lower explosive limits of combustible component j (ppmv)

n = number of combustible components in mixture

For the example case,

$$\sum_{i=1}^{n} x_i = (1,000 + 1,000) \times 10^{-6} \tag{3.10}$$
$$= 2,000 \times 10^{-6}$$

From standard references [14] or from Appendix 3A,

LEL_{Bz} = 14,000 ppmv for benzene
LEL_{MC} = 82,500 ppmv for methyl chloride

$$LEL_{mix} = \left[\frac{1,000}{2,000 \times 14,000} + \frac{1,000}{2,000 \times 82,500}\right]^{-1} \tag{3.11}$$
$$= 23,938 \text{ ppmv}$$

$$\% \, LEL_{mix} = \frac{\text{total combustible conc. in mixture}}{LEL_{mix}} \times 100 \tag{3.12}$$

$$= \frac{2,000}{23,938} \times 100 = 8.4\% \tag{3.13}$$

The percent LEL of the mixture is therefore 8.4%. Because this is well below 25%, no dilution air is needed in this example. If the mixture had been above 25% LEL, sufficient dilution air would have been needed to bring the concentration of the mixture to less than 25% to satisfy fire insurance regulations.

Step 4 - Calculate the volumetric heat of combustion of the waste gas stream, $(-\Delta h_{c_w})$, Btu/scf The energy content of the gas stream, expressed in terms of the heat of combustion, is calculated as follows:

$$(-\Delta h_{c_w}) = \sum_{i=1}^{n} (-\Delta h_{c_i}) x_i \tag{3.14}$$

where

$(-\Delta h_{c_w})$ = heat of combustion of the waste stream (Btu/scf)

$(-\Delta h_{c_i})$ = volumetric heat of combustion of component i at 25°C(Btu/scf)

x_i = volume fraction of component i in the waste gas

n = number of combustible components in the waste gas

The heat of combustion that should be used in these calculations is the "lower" heat of combustion, *i.e.*, with gaseous water, rather than liquid water, as a reaction product since water leaves the incinerator in the vapor state. From Appendix 3A or standard references [14,15] with appropriate conversion of units, the volumetric heat of combustion at 25°C for the two components is calculated to be as follows:

$$\begin{aligned}(-\Delta h_{c_{Bz}}) &= 3{,}475 \text{ Btu/scf for benzene}\\ (-\Delta h_{c_{MC}}) &= 705 \text{ Btu/scf for methyl chloride}\end{aligned}$$

The compositions specified earlier as ppmv are converted to volume fractions as follows:

$$\begin{aligned}x_{Bz} &= 1{,}000 \text{ ppmv} \times 10^{-6} = 10^{-3} \text{ for benzene}\\ x_{MC} &= 1{,}000 \text{ ppmv} \times 10^{-6} = 10^{-3} \text{ for methyl chloride}\end{aligned}$$

Using these values of heat of combustion and composition, the heat of combustion of the waste gas stream per *standard cubic foot* of incoming gas is

$$\begin{aligned}(-\Delta h_{c_w}) &= (3{,}475)\left(10^{-3}\right) + (705)\left(10^{-3}\right) \qquad (3.15)\\ &= 4.18 \text{ Btu/scf}\end{aligned}$$

Assuming the waste gas is principally air, with a molecular weight of 28.97 and a corresponding density of 0.0739 lb/scf, the heat of combustion per *pound* of incoming waste gas is

$$(-\Delta h_{c_w}) = 56.6 \text{ Btu/lb}$$

The negative heat of combustion, by convention, denotes an exothermic reaction. Also by convention, if one refers to heat of reaction rather than heat of combustion, then a positive value denotes an exothermic reaction.

Empirically, it has been found that 50 Btu/scf roughly corresponds to the LEL of organic/air mixtures. Insurance codes require a value below 25% LEL, which corresponds to about 13 Btu/scf. However, if LEL sensors and monitors are installed, one can incinerate a waste gas with a combustible organic content between 25 and 50% LEL, which corresponds to 13 to 25 Btu/scf.

For catalytic applications the heat of combustion must normally be less than 10 Btu/scf (for VOCs in air) to avoid excessively high temperatures in the catalyst bed. This is, of course, only an approximate guideline and may vary from system to system.

After Step 4, determination of $(-\Delta h_{c_w})$, the design procedure for thermal and catalytic incinerators is discussed separately, beginning with Step 5 for each type of incinerator.

3.4.2 Steps Specific to Thermal Units

Figure 3.1 shows a generic thermal incinerator with the appropriate streams labeled.

Step 5t - Establish the temperature at which the incinerator will operate As mentioned in Section 3.2.1, both the reactor temperature and residence time of the waste gas in the reactor determine the level of VOC destruction. In general, state and local regulations specify the required level of destruction that the customer must meet. In this example a destruction efficiency of 98 percent is specified. Studies by Mascone [2,3,4] show that this destruction efficiency can be met in a thermal incinerator operated at a temperature, T_{f_i}, of 1,600°F and a residence time of 0.75 second. (Note: This higher efficiency level is the *minimum* achievable by any new properly designed and operated incinerator. Many incinerators can achieve destruction efficiencies of 99% or higher.)

Step 6t - Calculate the waste gas temperature at the exit of the preheater The extent of heat exchange to be carried out in the preheater is the result of a technical and economic optimization procedure that is not illustrated in this example. As the VOC stream temperature leaving the heat exchanger, T_{w_o}, increases, the auxiliary fuel requirement decreases, but at the expense of a larger heat exchanger. However, there are several important limits on T_{w_o}. First, T_{w_o} must not be close to the ignition temperature of the organic-containing gas to prevent damaging temperature excursions inside the heat exchanger should the gas ignite. Second, for gases containing halogens, sulfur, and phosphorous (or other acid-forming atoms), the flue gas temperature after the heat exchanger, T_{f_o}, must not drop below the acid dew. Both limitations limit the amount of heat exchange and thus the maximum value of T_{w_o}. The calculation of the acid dew is not simple. It is recommended that vendor guidance be sought to ensure that the dew is not reached. Condensation of acid gases will result in corrosion of many of the metals used in heat exchangers. As an example, fuel sulfur contents of 1 to 2 percent can give acid dew points of about 200 to 270°F. Increasing the sulfur content to 4 percent can raise the dew to about 290°F. Chlorine and phosphorous have a much smaller effect on acid dew elevation.

With the following assumptions, one can estimate T_{w_o} using equation 3.2, the definition of fractional energy recovery for a heat exchanger.

- The fractional energy recovery is specified.
- The amount of auxiliary fuel, Q_{af}, and auxiliary combustion air, Q_a, are small relative to the waste gas, Q_w, so that the mass flow rates of gases, $\rho_w Q_w$ and $\rho_f Q_f$, on both sides of the preheater are approximately the same, or

$$\rho_w Q_w \approx \rho_f Q_f$$

- The heat capacities of the gases on both sides of the preheater are approximately the same, regardless of *composition*. This is true for waste streams which are dilute mixtures of organics in air, the properties of the streams changing only slightly on combustion.
- The *mean* heat capacities above the reference temperature of the gases on both sides of the preheater are approximately the same regardless of *temperature*.

With these assumptions, the equation for fractional energy recovery for a heat exchanger becomes

$$\text{Fractional Energy Recovery} = \frac{T_{w_o} - T_{w_i}}{T_{f_i} - T_{w_i}} \qquad (3.16)$$

For this example with a fractional energy recovery of 0.70, an incinerator operating temperature, T_{f_i}, of 1,600°F, and a waste gas inlet temperature, T_{w_i}, of 100°F, the waste gas temperature at the exit of the preheater becomes

$$T_{w_o} = 1,150^\circ\text{F}$$

The temperature of the exhaust gas, T_{f_o}, can be determined by an energy balance on the preheater, which, with the same assumptions as used in deriving Equation 3.16 regarding the mass flow rates and average heat capacities of the gases involved, results in the following equation:

$$T_{f_i} - T_{f_o} = T_{w_o} - T_{w_i}$$

i.e., the temperature rise in the waste gas is approximately equal to the temperature decrease in the flue gas with which it is exchanged. For this example, this results in

$$T_{f_o} = 550^\circ\text{F}$$

This value of T_{f_o} should be well above the acid dew of the flue gas stream.

It should be remembered that T_{w_o} should be well below the ignition temperature of the VOC stream to prevent unwanted temperature excursions in the preheater. This must be verified even if the stream is well below the LEL because flammability limits can be expanded by raising the reactant stream temperature. A sufficiently high preheat temperature, T_{w_o}, could initiate reaction (with heat release) in the preheater. This would ordinarily be detrimental to the materials of construction in the heat exchanger. The one exception is the thermal incinerator of the regenerable type described in Section 3.2 The 95 percent energy recovery obtainable in regenerable systems would result in this example in a T_{w_o} of 1,525°F. The significant reaction rate that would occur at this temperature in the ceramic packing of the heat exchanger/reactor is by design.

Step 7t - Calculate the auxiliary fuel requirement, Q_{af} Auxiliary fuel will almost invariably be needed for startup of the unit. However, at steady state, if the energy released by combustion of the organics present in the waste stream is sufficient to maintain the reactor temperature (1,600°F in the example), only a small amount of auxiliary fuel ($<$ 5% of the total energy input) is needed to stabilize the flame. In most cases, however, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature.

With the following assumptions, one can estimate Q_{af} using a mass and energy balance around the combustion chamber and following the principles discussed in Section 3.3, with reference to Figure 3.1.

- The reference temperature, T_{ref}, is taken as the inlet temperature of the auxiliary fuel, T_{af}.
- No auxiliary air, Q_a, is required.
- Energy losses, H_L, are assumed to be 10% of the total energy input to the incinerator above ambient conditions.[16,17] Thus, if the reference temperature is near ambient conditions,

$$H_L = 0.1\rho_{f_i} Q_{f_i} C_{p_{m_{f_i}}} (T_{f_i} - T_{ref}) \qquad (3.17)$$

- The heat capacities of the waste gases entering and leaving the combustion chamber are approximately the same, regardless of *composition.* This is true for waste streams which are dilute mixtures of organics in air, the properties of the streams changing only slightly on combustion.
- The *mean* heat capacities above the reference temperature of the waste gases entering and leaving the combustion chamber are approximately the same regardless of *temperature.* Thus the mean heat capacity for the waste gas stream entering or leaving the combustion chamber should be evaluated at the average of T_{w_o} and T_{f_i}. For air

this assumption introduces an error of, at most, 5% over the temperatures of interest.

With these assumptions, the mass and energy balance around the combustion chamber reduces to the following equation:

$$\rho_{af} Q_{af} = \frac{\rho_{w_o} Q_{w_o} [C_{p_{m_{air}}} (1.1 T_{f_i} - T_{w_o} - 0.1 T_{ref}) - (-\Delta h_{c_{w_o}})]}{(-\Delta h_{c_{af}}) - 1.1 C_{p_{m_{air}}} (T_{f_i} - T_{ref})} \qquad (3.18)$$

Input data for this equation are summarized below:

- The waste stream is essentially air so that

$$\rho_{w_o} = \rho_{w_i} = 0.0739 \text{ lb/scf}, \quad \text{air at 77°F, 1 atm.}$$

$C_{p_{m_{air}}} = 0.255$ Btu/lb °F, the mean heat capacity of air between 77°F and 1,375°F (the average temperature of the waste gas entering and leaving the combustion chamber)

- Other input data to Equation 3.18 include

$$\begin{aligned}
Q_{w_o} &= Q_{w_i} = 20,000 \text{ scfm} \\
(-\Delta h_{c_{af}}) &= 21,502 \text{ Btu/lb}, \quad \text{for methane} \\
T_{ref} = T_{af} &= 77°\text{F}, \quad \text{assume ambient conditions} \\
\rho_{af} &= 0.0408 \text{ lb/ft}^3, \quad \text{methane at 77°F, 1 atm.} \\
T_{f_i} &= 1,600°\text{F}, \quad \text{from Step 5t} \\
T_{w_o} &= 1,150°\text{F}, \quad \text{from Step 6t} \\
(-\Delta h_{c_{w_o}}) &= 56.6 \text{ Btu/lb}, \quad \text{from Step 4}
\end{aligned}$$

Substituting the above values into Equation 3.18 results in

$$Q_{af} = 167 \text{ scfm}$$

The values of the parameters in the energy balance are summarized in Table 3.5.

Table 3.5: Summary of Example Problem Variable Valuation

$T_{ref} = 77°F$

Stream	Subscript, j	ρ_j, lb/scf	Q_j, scfm	$C_{p_{m_j}}$, Btu/lb °F	T_j, °F
IN - Sensible Heat					
Auxiliary Air	a	na*	na*	na*	na*
Auxiliary Fuel	af	0.0408	167	**	77
Waste Gas	w_o	0.0739	20,000	0.255	1,150
OUT - Sensible Heat					
Waste Stream	f_i	0.0739	20,167	0.255	1,600

$(-\Delta h_c)$, waste gas = 56.6 Btu/lb
$(-\Delta h_c)$, auxiliary fuel = 21,502 Btu/lb
*Not applicable.
**Not used because reference temperature is taken equal to auxiliary fuel temperature.

It is instructive to examine the magnitude of the various terms in the energy balance around the combustor for the sample problem. This is done in Table 3.6. The energy balance shown does not quite add to zero due to round-off error and simplifying assumptions. Table 3.6 shows that the largest inlet term is the sensible heat of the incoming waste gas. The heat of combustion of the organics contained in the waste gas stream is somewhat smaller than that of the auxiliary methane because of the relatively small amount of organics in the waste gas stream. The largest term in the outlet stream is the sensible heat of the outgoing waste stream. The overall energy losses are based on an assumption, but are relatively small. Because the sensible heat contents of the entering and leaving waste stream are so large, it is apparent that energy recovery is an important factor in achieving energy efficiency. In fact, with zero energy recovery in the sample problem, the auxiliary fuel requirements would be 605 scfm, about four times the energy requirements based on 70% energy recovery.

Table 3.6: Terms in Energy Balance around Combustor – Example Problem

IN – OUT + GENERATION = 0

Stream	Subscript, j	Value, Btu/min
IN - Sensible Heat, $\rho_j Q_j C_{p_{m_j}}(T_i - T_{ref})$		
Auxiliary Air	a	0
Waste Gas	w_o	404,403
OUT - Sensible Heat, $\rho_j Q_j C_{p_{m_j}}(T_i - T_{ref})$		
Waste Stream	f_i	578,796
OUT - Losses		
10% of total energy input		57,880
GENERATION - Heat of Combustion, $\rho_j Q_j(-\Delta h_{c_j})$		
Waste Gas	w_o	83,655
Auxiliary Fuel	af	146,506

Step 8t - Verify that the auxiliary fuel requirement is sufficient to stabilize the burner flame Only a small amount of auxiliary fuel ($< 5\%$ of the total energy input) is needed to stabilize the burner flame. In general, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature. It is wise to verify that the auxiliary fuel requirement calculated in Step 7t is sufficient for stabilization. If it is insufficient, then a minimum amount of auxiliary fuel must be used, and the amount of energy recovery specified earlier must be reduced to avoid exceeding the specified temperature at which the incinerator will operate (Step 5t).

This check is made by calculating 5% of the total energy input to the incinerator and comparing it with the auxiliary fuel energy input. The total energy input is given as follows:

$$\text{Total Energy Input} = \rho_{f_i} Q_{f_i} C_{p_{m_{f_i}}} (T_{f_i} - T_{ref}) \qquad (3.19)$$

$$\text{Auxiliary Fuel Energy Input} = \rho_{af} Q_{af} (-\Delta h_{c_{af}}) \qquad (3.20)$$

The auxiliary fuel used in the design, Q_{af}, should be the larger of 5% of the total energy input (28,900 Btu/min.) and the auxiliary fuel energy input (146,500 Btu/min.). The auxiliary fuel used easily meets this criterion.

Step 9t - Calculate the total volumetric flow rate of gas through the incinerator, Q_{f_i} The total volumetric flow rate of gas leaving the incinerator is referred to as the flue gas flow rate, Q_{f_i}, and is the gas rate on which the incinerator sizing and cost correlations are based. The flue gas flow rate measured at the standard conditions of 77°F and 1 atmosphere, where the increase in volumetric throughput due to an increase in the number of moles of gas as a result of combustion is neglected, is the sum of the inlet streams to the incinerator.

$$\begin{aligned} Q_{f_i} &= Q_{w_o} + Q_a + Q_{af} \\ &= 20,000 + 0 + 167 \\ &= 20,167 \text{ scfm} \end{aligned}$$

This result conforms with the assumptions stated in Step 6t, *i.e.*, the mass (and volume) flow rates on both sides of the preheater are approximately equal.

3.4.3 Steps Specific to Catalytic Units

Figure 3.3 shows a generic catalytic incinerator with the appropriate streams labeled. The approach used in the calculations on the catalytic incinerator is somewhat different than that used in the thermal incinerator. This difference arises because of additional constraints which are placed on the catalytic incinerator. These constraints are as follows:

- The desired catalyst bed outlet temperature is typically 700 to 900 °F. The maximum temperature to which the catalyst bed can be exposed continuously is limited to about 1,200 °F. Therefore, the combustible content of the waste gas is limited, and the amount of heat exchange that occurs in the primary heat exchanger may be limited.
- The inlet temperature to the catalyst bed itself must be above the catalytic ignition temperature required to give the desired destruction efficiency in the incinerator. Therefore, the combustible content of the waste gas is further limited to that which, when combusted, will raise the temperature in the catalyst bed no more than the ΔT between the required reactor bed inlet temperature and the desired reactor bed outlet temperature.
- Auxiliary fuel, in combination with the preheat from the primary heat exchanger, is used to preheat the waste gas to the reactor inlet temperature. A minimum amount of auxiliary fuel (< 5% of the total energy input) must be used to stabilize the flame in the preheat combustion chamber. This has the effect of further limiting the combustible content of the waste gas stream and the amount of heat exchange permissible in the primary heat exchanger.

The steps outlined below represent one approach to recognizing these constraints and incorporating them into the calculation procedures.

Step 5c - Establish the desired outlet temperature of the catalyst bed, T_{fi} The energy released by the oxidation of the VOCs in the catalyst bed will raise the temperature of the gases by an amount, ΔT, as the gases pass through the catalyst bed. An outlet temperature from the catalyst, and thus from the reactor, must be specified that will ensure the desired level of destruction of the VOC stream. As in thermal incinerators, this value varies from compound to compound and also from catalyst to catalyst. Final design of the incinerator should be done by firms with experience in incinerator design. Guidelines given by Combustion Engineering [12] indicate that values from 300 to 900°F result in destruction efficiencies between 90 and 95 percent. To prevent deactivation of the catalyst a maximum bed temperature of 1,200°F should not be exceeded. In the example problem the catalyst outlet temperature, T_{f_i}, is selected to be 900°F.

Step 6c - Calculate the waste gas temperature at the exit of the preheater (primary) heat exchanger The waste gas temperature at the exit of the primary heat exchanger is estimated in the same manner as for the thermal incinerator. The equation for fractional energy recovery, Equation 3.16, is used, with the same assumptions as used for the thermal incinerator. For the example problem with a fractional energy recovery of 0.70, a catalyst bed outlet temperature, T_{f_i}, of 900°F, and a waste gas inlet temperature, T_{w_i}, of 100°F, the gas temperature at the exit of the preheater becomes

$$T_{w_o} = 660°\mathrm{F}$$

The same considerations regarding the closeness of the temperature of the exhaust gas, T_{f_o}, to its dew point apply to the catalytic incinerator as they did to the thermal incinerator.

Step 7c - Calculate the auxiliary fuel requirement, Q_{af} The auxiliary fuel requirement, Q_{af}, is calculated by making mass and energy balances around the preheater combustion chamber and the catalyst chamber. The auxiliary fuel requirement calculated in this manner must be checked to insure that it falls within the constraints imposed by design considerations of the catalytic incinerator. These constraints are as follows:

- The auxiliary fuel requirement must be positive. A negative fuel requirement indicates that the heat of combustion of the waste gas,

$(-\Delta h_c)$, is too high for the fractional energy recovery in the primary heat exchanger that was selected.

- The auxiliary fuel amount must be high enough to provide a stable flame in the preheater combustion chamber (See Step 8c below).

An energy balance around the preheater combustion chamber and the catalyst chamber, taken together, results in Equation 3.18, the same equation used in the thermal incinerator calculations. The input data for Equation 3.18 for the catalytic incinerator example problem are summarized below:

- The waste stream is essentially air so that

$$\begin{aligned}
\rho_{w_o} &= \rho_{w_i} = 0.0739 \text{ lb/scf}, \quad \text{air at 77°F, 1 atm} \\
C_{p_{m_{air}}} &= 0.248 \text{ Btu/lb °F}, \quad \text{the mean heat capacity of air between 77°F and 780°F(the average of the preheater exit and catalyst bed outlet temperatures)}
\end{aligned}$$

- Other input data to Equation 3.18 include

$$\begin{aligned}
Q_{w_o} &= Q_{w_i} = 20{,}000 \text{ scfm} \\
(-\Delta h_{c_{af}}) &= 21{,}502 \text{ Btu/lb}, \quad \text{for methane} \\
T_{af} &= 77°\text{F}, \quad \text{assume ambient conditions} \\
\rho_{af} &= 0.0408 \text{ lb/ft}^3, \quad \text{methane at 77°F, 1 atm} \\
T_{f_i} &= 900°\text{F}, \quad \text{from Step 5c} \\
T_{w_o} &= 660°\text{F}, \quad \text{from Step 6c} \\
(-\Delta h_{c_w}) &= 56.6 \text{ Btu/lb}, \quad \text{from Step 4}
\end{aligned}$$

Substituting the above values into Equation 3.18 results in

$$Q_{af} = 40 \text{ scfm}$$

If the outlet temperature of the catalyst bed, T_{f_i}, is 800°F, then Q_{af} decreases to -6.7 scfm. In other words, no auxiliary fuel would, theoretically, be required at this bed temperature. However, as discussed above in

Step 8t, a certain quantity of auxiliary fuel would be required to maintain burner stability.

At 70% energy recovery and 900°F outlet catalyst bed temperature, a waste gas with a heat of combustion, $(-\Delta h_{c_{wg}})$, of about 79.9 Btu/lb would cause the auxiliary fuel requirement, Q_{af}, to become negative, indicating the catalyst bed would exceed 900°F. At 70% energy recovery and 800°F outlet catalyst bed temperature, this same result occurs with a $(-\Delta h_{c_{wg}})$ of 52.7 Btu/lb. Both of these heats of combustion are relatively low for typical waste gases. These results are, of course, dependent on the assumption of energy losses from the combustion chamber. The lower the energy losses, the lower the allowable waste gas heat of combustion before overheating occurs in the catalyst bed.

Step 8c - Verify that the auxiliary fuel requirement is sufficient to stabilize the burner flame Only a small amount of auxiliary fuel (< 5% of the total energy input) is needed to stabilize the burner flame. In general, more fuel than just this stabilizing fuel will be required to maintain the reactor temperature. It is wise to verify that the auxiliary fuel requirement calculated in Step 7c is sufficient for stabilization. If it is insufficient, then a minimum amount of auxiliary fuel must be used and the amount of energy recovery specified earlier must be reduced to avoid exceeding the specified temperature at which the incinerator will operate (Step 5c).

This check is made in the same manner as that in Step 8t of the thermal incinerator calculation. The results of this check indicate that the auxiliary fuel requirement is more than sufficient to stabilize the burner flame.

Step 9c - Estimate the inlet temperature to the catalyst bed, T_{r_i} The inlet temperature to the catalyst bed must be calculated to ensure that the inlet temperature is above that necessary to ignite the combustible organic compounds in the catalyst that was selected for use.

The inlet temperature to the catalyst bed, T_{r_i}, should be such that, when the temperature rise through the catalyst bed, ΔT, is added to it, the resulting temperature is T_{f_i}, 900°F. Thus,

$$\Delta T = T_{f_i} - T_{r_i} \tag{3.21}$$

The value of ΔT is determined by an energy balance around the preheater portion of the combustor. The preheater is required to heat the gases up to the catalyst bed inlet temperature using auxiliary fuel.[1] This energy balance is made with the assumptions made earlier in deriving Equation 3.18 and further assuming that only auxiliary fuel is combusted in the preheater portion. The resulting equation is very similar to Equation 3.18 except that (1) the terms with an f_i subscript become terms with r_i subscripts to denote a catalytic *r*eactor *i*nlet stream rather than a combustor outlet (*f*lue gas *i*nlet to the primary heat exchanger) and (2) the term for combustion of the waste gas organics does not appear. The resulting equation is as follows:

$$\rho_{af}Q_{af} = \frac{\rho_{w_o}Q_{w_o}[C_{p_{m_{air}}}(1.1T_{r_i} - T_{w_o} - 0.1T_{ref})]}{(-\Delta h_{c_{af}}) - 1.1C_{p_{m_{air}}}(T_{r_i} - T_{ref})} \tag{3.22}$$

This equation may be rearranged to solve for T_{r_i} explicitly. This produces an equation that is somewhat complex and non-intuitive.

$$T_{r_i} = \frac{\rho_{af}Q_{af}[(-\Delta h_{c_{af}}) + 1.1C_{p_{m_{air}}}T_{ref}] + \rho_{w_o}Q_{w_o}C_{p_{m_{air}}}(T_{w_o} + 0.1T_{ref})}{1.1C_{p_{m_{air}}}(\rho_{af}Q_{af} + \rho_{w_o}Q_{w_o})} \tag{3.23}$$

After substituting the example problem parameters into Equation 3.23, we obtain a value for T_{r_i} of 693°F. Based on ignition temperatures shown in Table 3.2, this reactor inlet temperature should be satisfactory. Prior to a more definitive design, the ignition temperatures for the specific chemicals should be verified.

The temperature rise across the catalyst bed is thus (900 − 693) or 207°F. These temperatures are somewhat sensitive to the assumption for energy losses from the combustor. The assumption for energy losses is perhaps somewhat conservative, *i.e.*, it causes a larger Q_{af} to be estimated than would a less conservative assumption, and becomes more conservative as the combustor size and insulation are increased.

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[1] At equilibrium, the temperature of the catalyst bed is maintained without requiring auxiliary fuel.

Step 10c - Calculate the total volumetric flow rate of gas through the incinerator, Q_{f_i} The total volumetric flow rate of gas leaving the incinerator is referred to as the flue gas flow rate, Q_{f_i}, and is the gas rate on which the incinerator sizing and cost correlations are based. The flue gas flow rate measured at the standard conditions of 77°F and 1 atmosphere, where the increase in volumetric throughput due to an increase in the number of moles of gas as a result of combustion is neglected, is the sum of the inlet streams to the incinerator.

$$\begin{aligned} Q_{f_i} &= Q_{w_o} + Q_a + Q_{af} \\ &= 20,000 + 0 + 40 \\ &= 20,040 \text{ scfm} \end{aligned}$$

Step 11c - Calculate the volume of catalyst in the catalyst bed If the volumetric flow rate of gas through the catalyst bed, Q_{f_i}, and the nominal residence time (reciprocal space velocity) in the catalyst bed are known, then the volume of catalyst can be estimated. There exists a complex set of relationships between the catalyst volume and geometry, overall pressure drop across the catalyst, conversion of the oxidizable components in the gas, gas temperature, and the reaction rate. These relationships are dependent on the catalyst and the type of compound being oxidized. It is beyond the scope if this *Manual* to discuss these relationships, even in an approximate way. For the purposes of cost estimation, the space velocity, ϕ in reciprocal time units, necessary to achieve the required level of destruction can be used to approximate the catalyst volume requirement. The space velocity is defined as

$$\phi = \frac{Q_{f_i}}{V_{cat}}$$

where

V_{cat} = Overall bulk volume of the catalyst bed, including interparticle voids (ft^3)

By petro-chemical industry convention, the space velocity is computed at the conditions of 60°F (not 77°F) and 1 atm. The volumetric flow rate,

Q_{f_i}, must be corrected to these conditions. The proper space velocity to achieve a desired level of conversion is based on experimental data for the system involved. For precious metal monolithic catalysts, the space velocity generally lies between 10,000 h^{-1} and 60,000 h^{-1}. (Base metal catalysts operate at lower space velocities, ranging from 5,000 to 15,000 h^{-1}.)[10]

For the example, using a space velocity of 30,000 h^{-1} or 500 $\min^{-1}$, and using Q_{f_i} at 60°F,

$$\begin{aligned} Q_{f_i} \text{ at } 60^\circ\text{F} &= 20,040\left(\frac{60+460}{77+460}\right) \\ &= 19,400 \text{ ft}^3/\text{min} \\ V_{cat} &= \frac{19,400 \text{ ft}^3/\text{min}}{500 \text{ min}^{-1}} \\ &= 39 \text{ ft}^3 \end{aligned}$$

There are a number of catalyst bed parameters, such as catalyst configuration and bed design, that are not significant for study type cost estimates. Accordingly, design of these factors is not discussed here.

3.5 Cost Analysis

This section presents procedures and data for estimating capital and annual costs for four types of incinerators:(1) thermal-recuperative, (2) thermal-regenerative, (3) fixed-bed catalytic, and (4) fluid-bed catalytic.

3.5.1 Estimating Total Capital Investment

Total capital investment, TCI, includes the equipment cost, EC, for the incinerator itself, the cost of auxiliary equipment (*e.g.*, ductwork), all direct and indirect installation costs, and costs for buildings, site preparation, off-site facilities, land, and working capital. However, the last five costs usually do not apply to incinerators. (See Chapter 2 of this *Manual* for a detailed description of the elements comprising the TCI.)

Table 3.7: Scope of Cost Correlations

Incinerator Type	Total (Flue) Gas Flowrate, scfm	Figure Number
Thermal – Recuperative	500[a]-50,000	3.4
Thermal – Regenerative	10,000-100,000	3.5
Fixed-Bed Catalytic	2,000-50,000	3.6
Fluid-Bed Catalytic	2,000-25,000	3.7

[a] Although Figure 3.4 covers the 1,000 to 50,000 scfm range, the correlation is valid for the 500 to 50,000 scfm range.

3.5.1.1 Equipment Costs, EC

As discussed in Section 3.2.3, the equipment costs, EC, given in this chapter apply to packaged incinerators, except for regenerative incinerators. For regenerative incinerators, the costs apply to field-erected units. The EC typically includes all flange-to-flange equipment needed to oxidize the waste gas, including the auxiliary burners, combustion chamber, catalyst, primary heat exchanger (except for the "zero heat recovery" cases), weathertight housing and insulation, fan, flow and temperature control systems, a short stack, and structural supports. Smaller units, *e.g.* typically less than 20,000 scfm, are typically preassembled and skid-mounted [18]. The various available incineration systems are presented in four groups delineated according to their similarity of design. These groups are outlined in Table 3.7. With the exception of regenerative thermal and fluid-bed catalytic incinerators, the maximum size for which costs are given is 50,000 scfm. Although larger units of each technology can be built, applications are rare at flow rates above 50,000 scfm. Regenerative thermal incinerator costs are provided for flow rates from 10,000 to 100,000 scfm. Fluid-bed catalytic incinerator costs are provided for flow rates from 2,000 to 25,000 scfm.

The cost curves are least-squares regressions of cost data provided by different vendors. It must be kept in mind that even for a given incineration technology, design and manufacturing procedures vary from vendor to vendor, so that costs may vary. As always, once the study estimate is completed, it is recommended that more than one vendor be solicited for a more detailed cost estimate.

The additional expense of acid gas clean-up or particulate control is not treated in this section. The equipment cost of a gas absorber to remove any acid gases formed in the incinerator can be quite large, sometimes exceeding the equipment cost of the incinerator itself even for simple packed tower scrubbers [19]. For more complex absorbers that include venturi scrubbers instead of, or in addition to, packed beds, the cost of the scrubber alone may be up to 4 times that of the incinerator [11]. These more complex absorbers are sometimes necessary when particulates, in addition to acid gases, must be removed from the flue gas.

Thermal Incinerators Among the thermal units, the direct flame (0% energy recovery) and recuperative systems are treated together because the various levels of energy recovery are achieved simply by adding heat exchanger surface area. Costs for these units were provided by several vendors [12,20,21]. The EC of these units are given as a function of *total* volumetric throughput, Q_{tot}, in scfm. "Q_{tot}" is the total volume of the gaseous compounds exiting the combustion chamber; it is identical to the term, "Q_{f_i}", used in Figures 3.1 and 3.3. This includes the combustion products, nitrogen, unburned fuel and organics, and other constituents. (See Figure 3.4). Note that costs are given free on board (F.O.B.) in April 1988 dollars. Based on a least-squares regression analysis, a log-log relationship between throughput and EC was found for a given level of energy recovery (HR) over the flow rate range from 500 to 50,000 scfm. These relationships are as follows:

$$EC = 10294Q_{tot}^{0.2355} \qquad \text{HR} = 0\ \% \tag{3.24}$$
$$EC = 13149Q_{tot}^{0.2609} \qquad \text{HR} = 35\ \% \tag{3.25}$$
$$EC = 17056Q_{tot}^{0.2502} \qquad \text{HR} = 50\ \% \tag{3.26}$$
$$EC = 21342Q_{tot}^{0.2500} \qquad \text{HR} = 70\ \% \tag{3.27}$$

The regenerative (or excess enthalpy) systems provide up to 95 percent heat recovery at the expense of higher capital costs. Their unique design [22,23], which combines the heat exchanger and reactor, is substantially different from traditional thermal units and is therefore treated separately in Figure 3.5. The ECs of these systems are given as an approximately

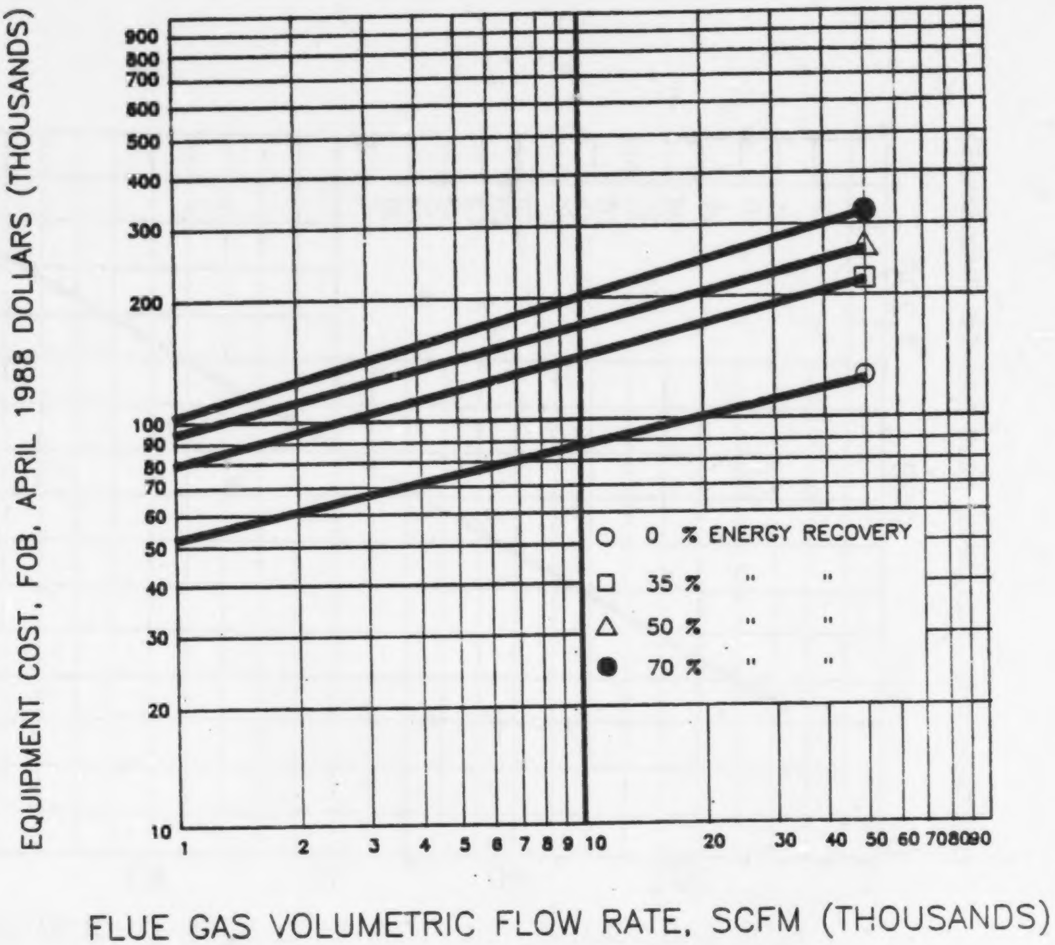

Figure 3.4: Equipment Costs of Thermal Incinerators, Recuperative

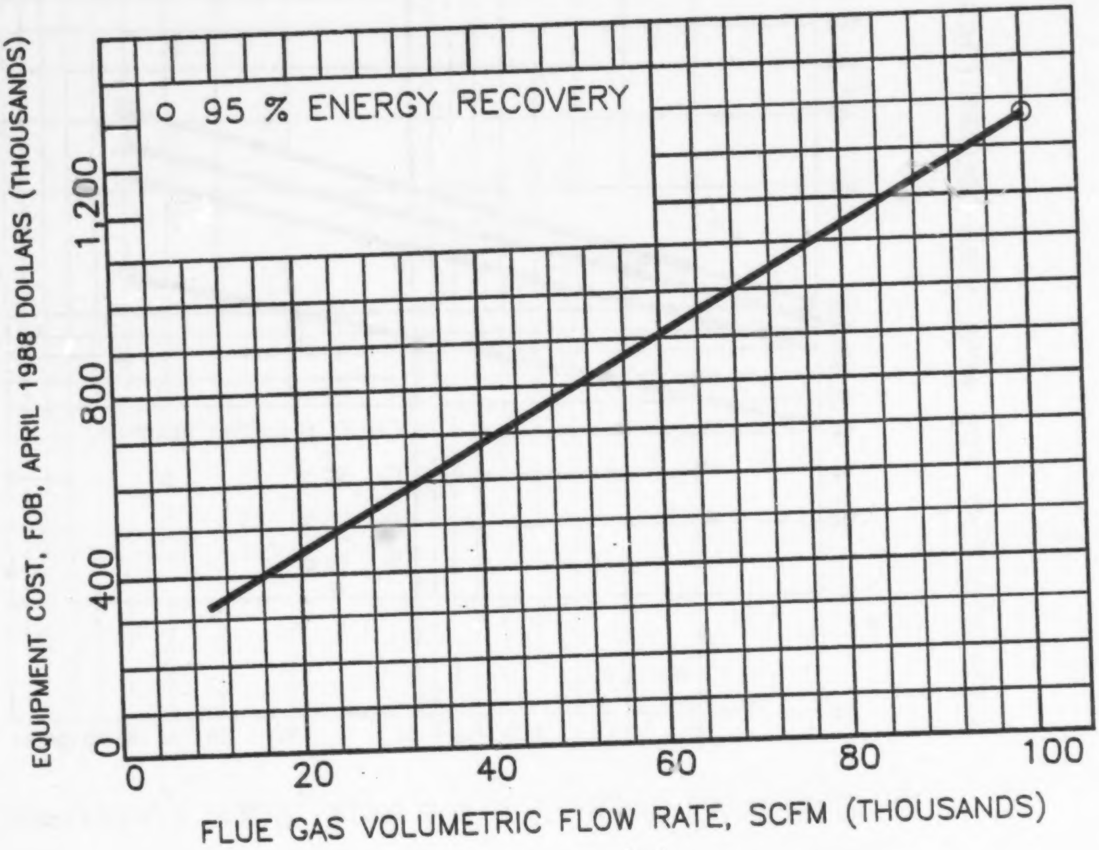

Figure 3.5: Equipment Costs of Thermal Incinerators, Regenerative

linear function of total flow rate over a 10,000 to 100,000 scfm range by the following equation:

$$EC = 2.204 \times 10^5 + 11.57\, Q_{tot} \tag{3.28}$$

Again, the higher capital costs of these units can be substantially offset by the substantial savings in auxiliary fuel costs.

Catalytic Incinerators The EC for a catalytic incinerator is a function of the type of catalyst contacting pattern used and the total gas flow rate, Q_{tot}, for a given level of energy recovery. There are three types of contacting configurations used in catalytic systems: fixed-bed, catalytic monolith, and fluid-bed. The EC for the first two are generally comparable and are given in Figure 3.6. The data provided by several vendors [12,20,21,24] exhibited curvilinear relationships with Q_{tot} for each of the energy recovery rates. Least squares regressions of the data yielded the following correlations for total flow rates between 2,000 and 50,000 scfm:

$$EC = 1105 Q_{tot}^{0.5471} \qquad \mathrm{HR} = 0\ \% \tag{3.29}$$

$$EC = 3623 Q_{tot}^{0.4189} \qquad \mathrm{HR} = 35\ \% \tag{3.30}$$

$$EC = 1215 Q_{tot}^{0.5575} \qquad \mathrm{HR} = 50\ \% \tag{3.31}$$

$$EC = 1443 Q_{tot}^{0.5527} \qquad \mathrm{HR} = 70\ \% \tag{3.32}$$

Fluid-bed catalytic incinerators afford certain advantages over fixed-bed catalyst units in that they tolerate waste streams with (1) higher heating values, (2) particulate contents, and (3) chlorinated species. For this enhanced flexibility of feed streams, a higher capital cost is incurred, as indicated by the EC shown in Figure 3.7. The data shown were provided by vendors [11,19] and exhibited a linear relationship over the range of flow rates from 2,000 to 25,000 scfm. They can be approximated by the following equations:

$$EC = 8.48 \times 10^4 + 13.2 Q_{tot} \qquad \mathrm{HR} = 0\ \% \tag{3.33}$$

$$EC = 8.84 \times 10^4 + 14.6 Q_{tot} \qquad \mathrm{HR} = 35\ \% \tag{3.34}$$

$$EC = 8.66 \times 10^4 + 15.8 Q_{tot} \qquad \mathrm{HR} = 50\ \% \tag{3.35}$$

$$EC = 8.39 \times 10^4 + 19.2 Q_{tot} \qquad \mathrm{HR} = 70\ \% \tag{3.36}$$

A comparison of the EC of thermal, catalytic fixed-bed, and catalytic fluid-bed systems with 50 percent energy recovery is shown in Figure 3.8.

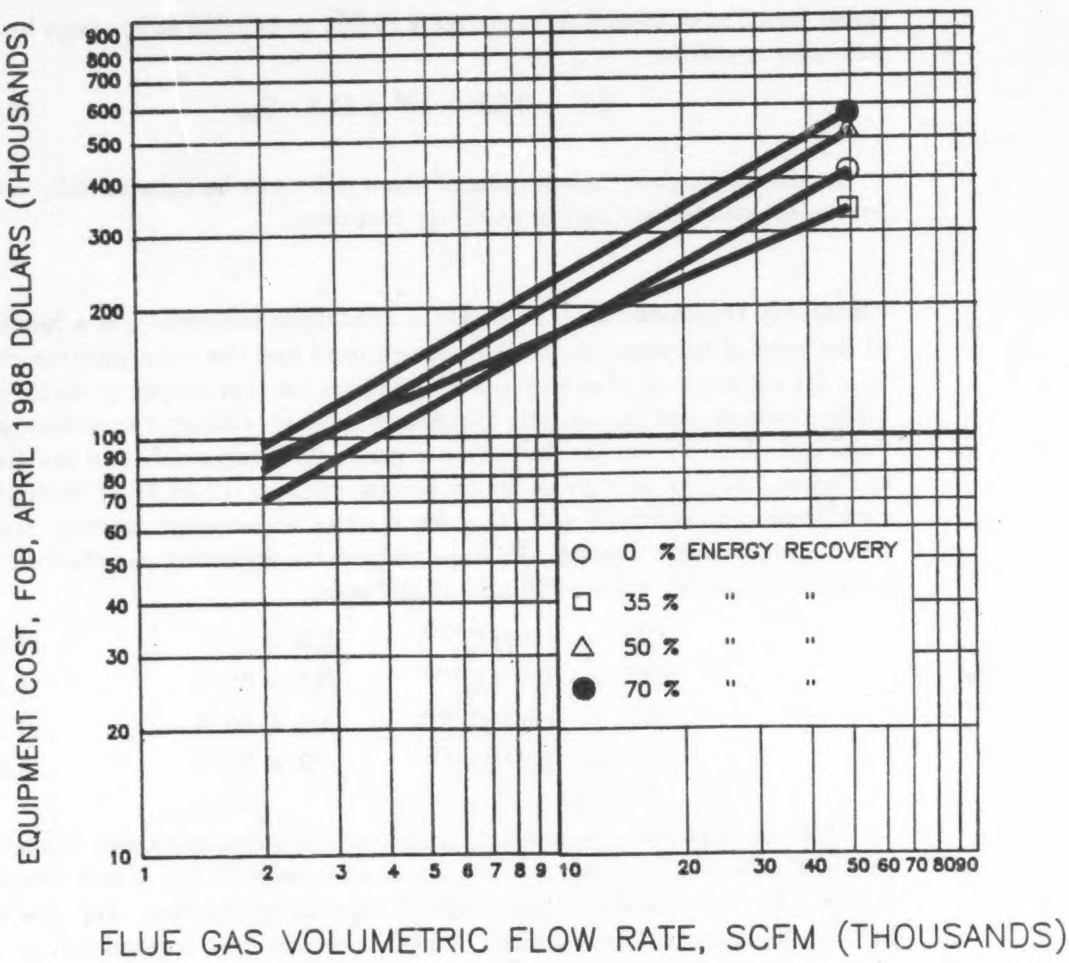

Figure 3.6: Equipment Costs of Catalytic Incinerators, Fixed-Bed

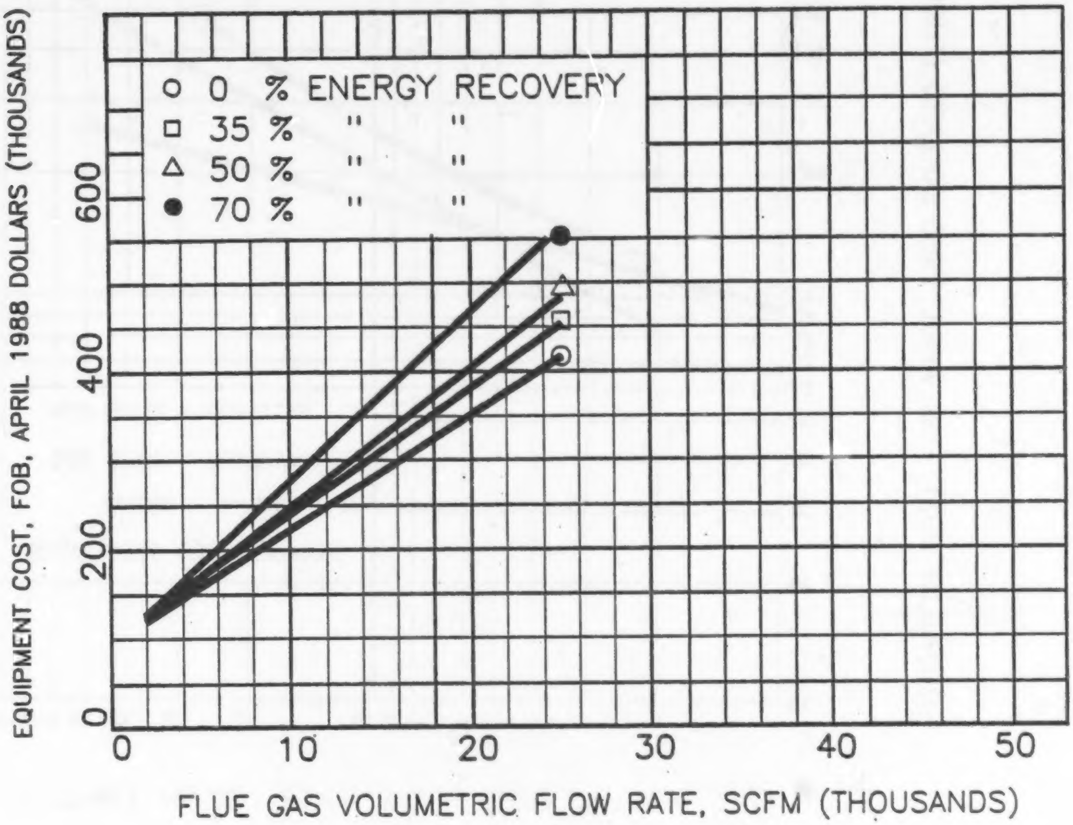

Figure 3.7: Equipment Costs of Catalytic Incinerators, Fluid-Bed

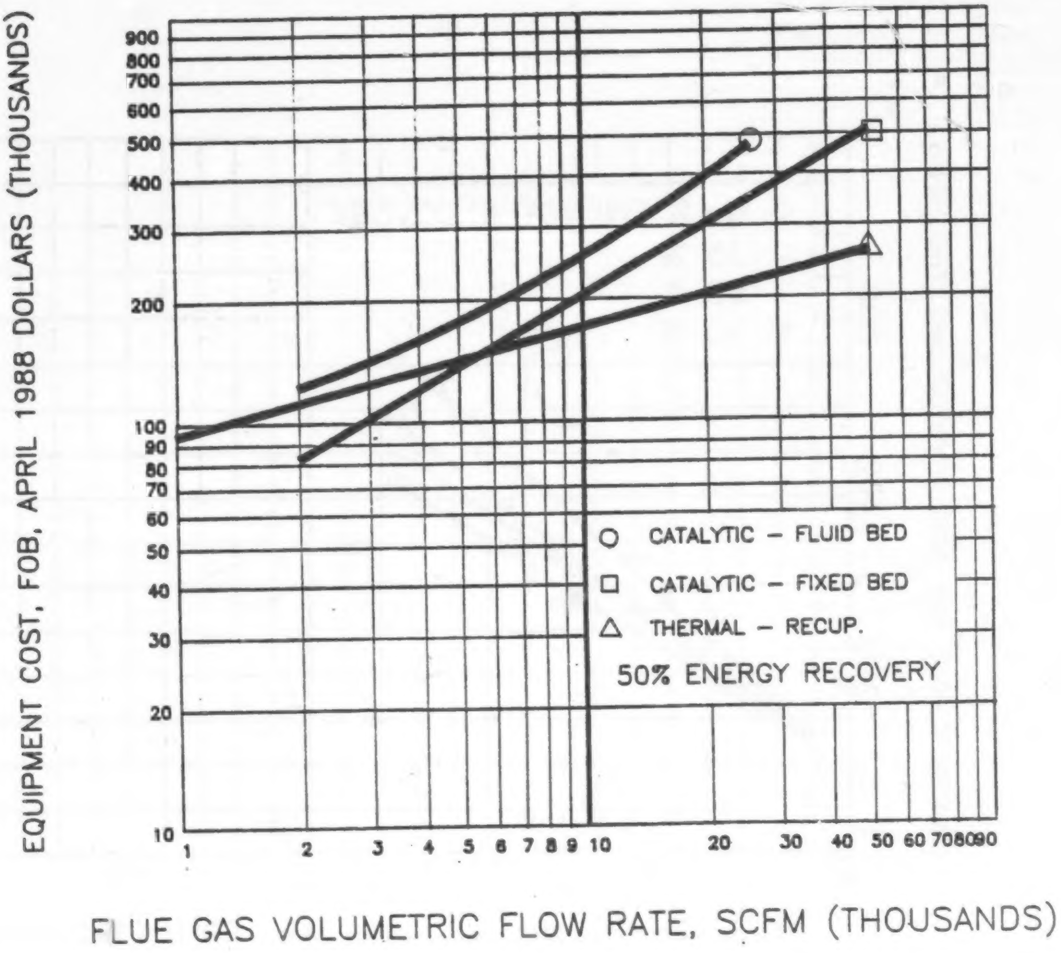

Figure 3.8: Equipment Cost Comparison of Incinerator Types

3.5.1.2 Installation Costs

As explained in Chapter 2, the purchased equipment cost, PEC, is calculated by taking the sum of the EC and the cost of auxiliary equipment (*e.g.*, ductwork), taxes, freight, and instrumentation. Average values of direct and indirect installation factors [25] to be applied to the PEC are given in Table 3.8 for both recuperative thermal and fixed- and fluid-bed catalytic incinerators.

Table 3.9 shows the itemized installation costs that are obtained when these installation factors are applied to the PECs for the example incinerators. Depending on the site conditions, the installation costs for a given incinerator could deviate significantly from costs generated by these average factors. Vatavuk and Neveril [25] provide some guidelines for adjusting the average installation factors to account for other-than-average installation conditions. For units handling total gas flow rates lower than 20,000 scfm the installation costs are minimal, amounting normally to only utility tie-ins (electrical and, if necessary, combustion or dilution air). The installation costs for these smaller incinerators would be 20 to 25 % of the PEC. Smaller units may be installed on the roofs of manufacturing buildings rather than at ground level. In such cases the installation factors could be as high as (or higher than) the factors shown in Table 3.8, even though the units would be "packaged".

3.5.2 Estimating Total Annual Cost

The total annual cost (TAC) is the sum of the direct and indirect annual costs. The TAC for both example systems is given in Table 3.10, along with suggested factors for calculating them.

3.5.2.1 Direct Annual Costs

Direct annual costs for incinerators include labor (operating and supervisory), maintenance (labor and materials), fuel, electricity, and (in catalytic units) replacement catalyst. For thermal and catalytic units, the fuel usage rate is calculated as shown in Sections 3.4.2 and 3.4.3, respectively, where

Table 3.8: Capital Cost Factors for Thermal and Catalytic Incinerators[a]

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
Incinerator (EC) + auxiliary equipment[b]	As estimated, A
Instrumentation[c]	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.08 B
Handling & erection	0.14 B
Electrical	0.04 B
Piping	0.02 B
Insulation for ductwork[d]	0.01 B
Painting	0.01 B
Direct installation cost	0.30 B
Site preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Cost, DC	1.30 B + SP + Bldg.
Indirect Costs (installation)	
Engineering	0.10 B
Construction and field expenses	0.05 B
Contractor fees	0.10 B
Start-up	0.02 B
Performance test	0.01 B
Contingencies	0.03 B
Total Indirect Cost, IC	0.31 B
Total Capital Investment = DC + IC	1.61 B + SP + Bldg.

[a] Reference [25].

[b] Ductwork and any other equipment normally not included with unit furnished by incinerator vendor.

[c] Instrumentation and controls often furnished with the incinerator, and thus often included in the EC.

[d] If ductwork dimensions have been established, cost may be estimated based on \$10 to \$12/ft^2 of surface for field application. Fan housings and stacks may also be insulated.

Table 3.9: Capital Costs for Thermal and Catalytic Incinerators Example Problem

Cost Item	Cost, $	
	Thermal-Recuperative	Fluid-Bed Catalytic
Direct Costs		
Purchased equipment costs		
Incinerator (EC)	$254,200	$468,200
Auxiliary equipment[a]	—	—
Sum = A	$254,200	$468,200
Instrumentation, 0.1A	25,400	46,800
Sales taxes, 0.03A	7,630	14,000
Freight, 0.05A	12,700	23,400
Purchased equipment cost, B	$300,000	$552,400
Direct installation costs		
Foundation and supports, 0.08B	24,000	44,200
Handling and erection, 0.14B	42,000	77,300
Electrical, 0.04B	12,000	22,100
Piping, 0.02B	6,000	11,000
Insulation (for ductwork), 0.01B	3,000	5,520
Painting, 0.01B	3,000	5,520
Direct installation cost	$90,000	$165,600
Site preparation[a]	—	—
Buildings[a]	—	—
Total Direct Cost	$390,000	$718,000
Indirect Costs (installation)		
Engineering, 0.10B	30,000	55,200
Construction and field expenses, 0.05B	15,000	27,600
Contractor fees, 0.10B	30,000	55,200
Start-up, 0.02B	6,000	11,000
Performance test, 0.01B	3,000	5,520
Contingencies, 0.03B	9,000	16,600
Total Indirect Cost	$93,000	$171,100
Total Capital Investment (rounded)	$483,000	$889,000

[a] None of these items is required.

Table 3.10: Annual Costs for Thermal and Catalytic Incinerators

Example Problem

Cost Item	Suggested Factor	Unit Cost[a]	Thermal	Fluid-Bed Catalytic
Direct Annual Costs[b], DC				
Operating Labor				
Operator	0.5 h/shift	$12.96/h	6,480	6,480
Supervisor	15% of operator	—	972	972
Operating materials	—			
Maintenance				
Labor	0.5 h/shift	$14.26/h	7,130	7,130
Material	100% of maint. labor	—	7,130	7,130
Catalyst replacement	100% of catalyst replaced ea 2 yr	$650/ft^3 for metal oxide	0	14,600
Utilities				
Natural Gas	—	$3.30/kft^3	264,500	63,400
Electricity	—	$0.059/kWh	35,000	42,300
Total DC			$321,200	$142,000
Indirect Annual Costs, IC				
Overhead	60% of sum of operating, supv., & maint. labor & maint. materials.	—	13,000	13,000
Administrative charges	2% TCI	—	9,660	17,800
Property taxes	1% TCI	—	4,830	8,900
Insurance	1% TCI	—	4,830	8,900
Capital recovery[c]	CRF [TCI - 1.08 (Cat. Cost)]	—	78,600	142,200
Total IC			$110,900	$190,800
Total Annual Cost (rounded)			$432,000	$352,000

[a] 1988 dollars.

[b] Assumes 8,000 h/yr.

[c] The capital recovery cost factor, CRF, is a function of the catalyst or equipment life (typically, 2 and 10 years, respectively) and the opportunity cost of the capital (*i.e.*, interest rate). For example, for a 10 year equipment life and a 10% interest rate, CRF = 0.1628.

Table 3.11: Typical Pressure Drop Across Selected Equipment

Equipment Type	Energy Recovery, %	Δ P, in. H_2O
Thermal Incinerators	0	4
Catalytic Fixed-bed Incinerators	0	6
Catalytic Fluid-bed Incinerators	0	6-10
Heat exchangers	35	4
" "	50	8
" "	70	15

natural gas (methane) is assumed to be the fuel. (Other fuels could be used for thermal units.)

The electricity costs are primarily associated with the fan needed to move the gas through the incinerator. The power (in kilowatts) needed to move a given volumetric flow rate of air (Q_{tot} per Sections 3.4.2 and 3.4.3) at a total flange-to-flange pressure drop of Δ P inches of water and combined motor/fan efficiency, ϵ, is adapted from Equation 2-7, as follows:

$$Power_{fan} = \frac{1.17 \times 10^{-4} Q_{tot} \Delta P}{\epsilon} \tag{3.37}$$

Fan efficiencies vary from 40 to 70 percent [15] while motor efficiencies are usually 90 percent.

The total pressure drop across an incinerator system depends on the number and types of equipment elements included in the system and on design considerations. The estimation of actual pressure drop requirements involves complex calculations based on the specific system's waste gas and flue gas conditions and equipment used. For the purposes of this section, however, the approximate values shown in Table 3.11 can be used.

For the example cases, we will assume 8,000 hours per year operation and a 60% efficiency for the fan and motor together. Using pressure drops of 4 and 8 inches of water, respectively, for the thermal and fluid-bed catalytic incinerators[2], and adding the pressure drop of 15 inches of water for 70%

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[2] A fluid-bed catalytic incinerator is used because the waste gas contains a chlorinated compound which would poison the catalyst in a fixed-bed incinerator.

heat recovery, the fan power requirements can be calculated as follows:

Thermal Incinerator

$$\begin{aligned} Power_{fan} &= \frac{1.17 \times 10^{-4}(20,000 \text{ scfm})(19 \text{ inches water})}{0.60} \\ &= 74.1 \text{ kW} \end{aligned}$$

Catalytic Incinerator

$$\begin{aligned} Power_{fan} &= \frac{1.17 \times 10^{-4}(20,000 \text{ scfm})(23 \text{ inches water})}{0.60} \\ &= 89.7 \text{ kW} \end{aligned}$$

The annual electricity costs would be the products of these usages, the annual operating hours, and the electricity cost (\$/kWh), or:

$$\begin{aligned} \text{Electricity Cost (Thermal)} &= 74.1 \text{ kW} \times 8,000 \text{ hours/yr} \times \$0.059/\text{kWh} \\ &= \$35,000 \text{ per yr} \\ \text{Electricity Cost (Catalytic)} &= 89.7 \text{ kW} \times 8,000 \text{ hours/yr} \times \$0.059/\text{kWh} \\ &= \$42,300 \text{ per yr} \end{aligned}$$

The catalyst replacement costs and scheduling are highly variable and depend on the nature of the catalyst, the amount of "poisons" and particulates in the gas stream (including the auxiliary fuel), the temperature history of the catalyst, and the design of the unit. It is impossible to predict the costs in a general sense. However, noble metal monolith catalysts operating on pure hydrocarbon gases in air will last longer than fluid-bed base metal catalysts operating on chlorinated hydrocarbons in air. Noble metal catalysts are also more expensive than base metal oxide catalysts. The catalyst life for many field units is from 1 to 4 years. The cost, in April 1988 dollars, of the replacement catalyst must be obtained from the vendor, but it may be estimated at \$3,000/ft^3 for noble metal catalysts and \$650/ft^3 for base metal oxide catalysts. For the example case, the catalyst is a base metal oxide because the waste gas contains a chlorinated compound. We will assume a two year catalyst life. Knowing that the catalyst volume is 39 ft^3 (Section 3.4.3) and using a cost of \$650/ft^3 and a capital recovery factor of 0.5762 (2-year life at a 10% interest rate), the annual expense for catalyst replacement is

$$\begin{aligned}\text{Annual Catalyst Replacement Cost} &= 39\ \text{ft}^3 \times 650\frac{\$}{\text{ft}^3} \times 0.5762\\ &= \$14,600 \text{ per year}\end{aligned}$$

To calculate the fuel or electricity annual cost, multiply the fuel usage rate (scfm) or the electricity usage rate (kW) by the total hours per year of operation (*e.g.*, 333 d/yr × 24 h/d = 8,000 h/yr) and by the appropriate unit cost (*e.g.*, $/scfm for fuel and $/kWh for electricity).

For the example cases, the fuel costs can be calculated from the fuel usage rates and the natural gas unit cost of $0.00330 /scf. For the thermal incinerator example, the annual fuel cost is calculated as follows:

$$\begin{aligned}\begin{matrix}\text{Annual Fuel}\\ \text{Cost, Thermal}\end{matrix} &= 0.00330\frac{\$}{\text{scf}} \times 167\frac{\text{scf}}{\text{min}} \times 60\frac{\text{min}}{\text{hr}} \times 8,000\frac{\text{hr}}{\text{yr}}\\ &= \$264,500 \text{ per yr}\end{aligned}$$

For the catalytic incinerator example. the annual fuel cost is found similarly:

$$\text{Annual Fuel Cost, Catalytic} = \$63,400 \text{ per year}$$

Operating and maintenance labor are estimated as 0.5 hours per 8-hour shift each, supervisory labor at 15 % of operating labor, and maintenance material as 100 % of maintenance labor.

3.5.2.2 Indirect Annual Costs

The indirect (fixed) annual costs include capital recovery, overhead, and property taxes, insurance, and administrative (G&A) charges. The last three of these can be estimated at 1%, 1%, and 2% of the total capital investment, respectively. The system capital recovery cost is based on an estimated 10-year equipment life. (See Section 2 for a thorough discussion of the capital recovery cost and the variables that determine it.) The system capital recovery cost is the product of the system capital recovery factor (CRF) and the total capital investment (TCI) less the purchased cost of the catalyst ($C_{cat} \times 1.08$ where the 1.08 is for freight and sales tax). These values calculated for the example cases are given in Table 3.10.

3.5.3 Cost Comparison for Example Case

The example VOC stream defined in Section 3.4.1 serves to illustrate some typical characteristics of thermal and catalytic systems. The total annual costs shown in Table 3.10 show that the catalytic system's auxiliary fuel costs are significantly lower than those of the thermal unit. The disparity is enough to offset the higher capital costs of the catalytic incinerator over the assumed 10-year lifetime of the units. Two factors that should be noted in the comparison of these two systems are (1) the 98 percent level of destruction met by the thermal incinerator may be difficult to reach by the catalytic system (this may be important in some cases), and (2) the example waste stream is of particularly low heating value (4 Btu/scf) which favors the catalytic system due to the lower auxiliary fuel requirements.

3.6 Acknowledgements

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- Energy Development Associates (EDA) (Itasca, IL)
- Pillar Technologies, Inc. (Hartland, WI)
- Huntington Energy Systems, Inc. (Union, NJ)

- **Regenerative Environmental Equipment Co. (REECO) (Morris Plains, NJ)**

- **Englehard Corp. (Edison, NJ)**

Appendix 3A

Properties of Selected Compounds

Table 3.12: Limits of Flammability of Combustible Organic Compounds in Air at Atmospheric Pressure, Room Temperature[*]

Compound	Molecular Weight	LEL[a], vol. %	UEL[b], vol. %
Methane	16.04	5.00	15.00
Ethane	30.07	3.00	12.50
Propane	44.09	2.12	9.35
Butane	58.12	1.86	8.41
Pentane	72.15	1.40	7.80
Hexane	86.17	1.18	7.40
Octane	114.23	0.95	
Nonane	128.25	0.83	
Decane	142.28	0.77	
Ethylene	28.05	2.75	28.60
Propylene	42.08	2.00	11.10
Acetylene	26.04	2.50	80.00
Cyclohexane	84.16	1.26	7.75
Benzene	78.11	1.40	7.10
Toluene	92.13	1.27	6.75

[*] Reference [14]

[a] Lower Explosive Limit

[b] Upper Explosive Limit

Table 3.13: Molar Heat Capacities of Gases at Zero Pressure [*]

$$C_p = a + bT + cT^2 + dT^3 \; ; \quad \text{T in °K}$$

$$C_{p_m} = \frac{\int_{T_1}^{T_2} C_p dT}{(T_2 - T_1)}$$

C_p in calories/g-mole °K or Btu/lb-mole °R

Compound	a	b ×10^2	c ×10^5	d ×10^9	Temperature Range, °K
Methane	4.750	1.200	0.3030	–2.630	273-1500
Ethane	1.648	4.124	–1.530	1.740	273-1500
Propane	–0.966	7.279	–3.755	7.580	273-1500
Butane	0.945	8.873	–4.380	8.360	273-1500
Pentane	1.618	10.85	–5.365	10.10	273-1500
Hexane	1.657	13.19	–6.844	13.78	273-1500
Cyclopentane	–12.957	13.087	–7.447	16.41	273-1500
Cyclohexane	–15.935	16.454	–9.203	19.27	273-1500
Benzene	–8.650	11.578	–7.540	18.54	273-1500
Toluene	–8.213	13.357	–8.230	19.20	273-1500
Nitrogen	6.903	–0.03753	0.1930	–0.6861	273-1800
Oxygen	6.085	0.3631	–0.1709	0.3133	273-1800
Air	6.713	0.04697	0.1147	–0.4696	273-1800
Carbon dioxide	5.316	1.4285	–0.8362	1.784	273-1800

[*] Reference [26]

Table 3.14: Heats of Combustion of Selected Gaseous Organic Compounds, $-\Delta h_c$, at 25°C and constant pressure to form gaseous water and carbon dioxide.*

Compound	Molecular Weight	$-\Delta h_c$	
		cal/g.	Btu/lb
Methane	16.04	11,953.6	21,502
Ethane	30.07	11,349.6	20,416
Propane	44.09	11,079.2	19,929
Butane	58.12	10,932.3	19,665
Pentane	72.15	10,839.7	19,499
Hexane	86.17	10,780.0	19,391
Octane	114.23	10,737.2	19,256
Nonane	128.25	10,680.0	19,211
Decane	142.28	10,659.7	19,175
Ethylene	28.05	11,271.7	20,276
Propylene	42.08	10,942.3	19,683
Cyclopentane	70.13	10,563.1	19,001
Cyclohexane	84.16	10,476.7	18,846
Benzene	78.11	9,698.4	17,446
Toluene	92.13	9,784.7	17,601

*Reference [15]

References

[1] *Prudent Practices for Disposal of Chemicals from Laboratories*, National Academy Press, Washington, D.C., 1983.

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Chapter 4

CARBON ADSORBERS

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Contents

4.1 Process Description 4-3

  4.1.1 Introduction 4-3

  4.1.2 Types of Adsorbers 4-4

    4.1.2.1 Fixed-bed Units 4-4

    4.1.2.2 Cannister Units 4-7

  4.1.3 Adsorption Theory 4-8

4.2 Design Procedure 4-14

  4.2.1 Sizing Parameters 4-14

  4.2.2 Determining Adsorption and Desorption Times 4-16

  4.2.3 Estimating Carbon Requirement 4-18

    4.2.3.1 Overview of Carbon Estimation Procedures . 4-18

    4.2.3.2 Carbon Estimation Procedure Used in *Manual* 4-18

4.3 Estimating Total Capital Investment 4-20

  4.3.1 Fixed-Bed Systems 4-20

    4.3.1.1 Carbon Cost 4-21

    4.3.1.2 Vessel Cost 4-21

    4.3.1.3 Total Purchased Cost 4-24

    4.3.1.4 Total Capital Investment 4-25

  4.3.2 Cannister Systems 4-25

4.4 Estimating Total Annual Cost 4-27

4.4.1 Direct Annual Costs 4-28
    4.4.1.1 Steam 4-28
    4.4.1.2 Cooling Water 4-28
    4.4.1.3 Electricity 4-29
    4.4.1.4 Carbon Replacement 4-32
    4.4.1.5 Solid Waste disposal 4-32
    4.4.1.6 Operating and Supervisory Labor 4-33
    4.4.1.7 Maintenance Labor and Materials 4-33
4.4.2 Indirect Annual Costs 4-33
4.4.3 Recovery Credits 4-34
4.4.4 Total Annual Cost 4-35
4.4.5 Example Problem 4-35
References 4-43

4.1 Process Description

4.1.1 Introduction

In air pollution control, adsorption is employed to remove volatile organic compounds (VOC's) from low to medium concentration gas streams, when a stringent outlet concentration must be met and/or recovery of the VOC is desired. Adsorption itself is a phenomenon where gas molecules passing through a bed of solid particles are selectively held there by attractive forces which are weaker and less specific than those of chemical bonds. During adsorption, a gas molecule migrates from the gas stream to the surface of the solid where it is held by physical attraction releasing energy—the "heat of adsorption", which approximately equals the heat of condensation.

Adsorptive capacity of the solid for the gas tends to increase with the gas phase concentration, molecular weight, diffusivity, polarity, and boiling point.

Some gases form actual chemical bonds with the adsorbent surface groups. This phenomenon is termed "chemisorption".

Most gases ("adsorbates") can be removed ("desorbed") from the adsorbent by heating to a sufficiently high temperature, usually via steam or (increasingly) hot combustion gases, or by reducing the pressure to a sufficiently low value (vacuum desorption). The physically adsorbed species in the smallest pores of the solid and the chemisorbed species may require rather high temperatures to be removed, and for all practical purposes cannot be desorbed during regeneration. For example, approximately 3 to 5 percent of organics adsorbed on virgin activated carbon is either chemisorbed or very strongly physically adsorbed and, for all intents, cannot be desorbed during regeneration.[1]

Adsorbents in large scale use include activated carbon, silica gel, activated alumina, synthetic zeolites, fuller's earth, and other clays. This chapter is oriented toward the use of activated carbon, a commonly used adsorbent for VOCs.

4.1.2 Types of Adsorbers

Five types of adsorption equipment are used in collecting gases: (1) fixed regenerable beds; (2) disposable/rechargeable cannisters; (3) traveling bed adsorbers; (4) fluid bed adsorbers; and (5) chromatographic baghouses.[2] Of these, the most commonly used in air pollution control are the fixed-bed and cannister types. This chapter addresses only fixed-bed and cannister units.

4.1.2.1 Fixed-bed Units

Fixed-bed units can be sized for controlling continuous, VOC-containing streams over a wide range of flow rates, ranging from several hundred to several hundred thousand cubic feet per minute (cfm). The VOC concentration of streams that can be treated by fixed-bed adsorbers can be as low as several parts per billion by volume (ppbv) in the case of some toxic chemicals or as high as 25% of the VOCs' lower explosive limit (LEL). (For most VOCs, the LEL ranges from 2500 to 10,000 ppmv.[3])

Fixed-bed adsorbers may be operated in either *intermittent* or *continuous* modes. In intermittent operation, the adsorber removes VOC for a specified time (the "adsorption time"), which corresponds to the time during which the controlled source is emitting VOC. After the adsorber and the source are shut down (*e.g.*, overnight), the unit begins the *desorption* cycle during which the captured VOC is removed from the carbon. This cycle, in turn, consists of three steps: (1) *regeneration* of the carbon by heating, generally by blowing steam through the bed in the direction opposite to the gas flow;[1] (2) *drying* of the bed, with compressed air or a fan; and (3) *cooling* the bed to its operating temperature via a fan. (In most designs, the same fan can be used both for bed drying and cooling.) At the end of the desorption cycle (which usually lasts 1 to $1\frac{1}{2}$ hours), the unit sits idle until the source starts up again.

In *continuous* operation a regenerated carbon bed is always available for adsorption, so that the controlled source can operate continuously without shut down. For example, two carbon beds can be provided: while one is adsorbing, the second is desorbing/idled. As each bed must be large enough to handle the entire gas flow while adsorbing, twice as much carbon must be provided than an *intermittent* system handling the same flow. If the desorption cycle is significantly shorter than the adsorption cycle, it may be more economical to have three, four, or even more beds operating in the system. This can reduce the amount of extra carbon capacity needed or provide some additional benefits, relative to maintaining a low VOC content in the effluent. (See Section 4.2 for a more thorough discussion of this.)

A typical two-bed, continuously operated adsorber system is shown in Figure 4.1. One of the two beds is adsorbing at all times, while the other is desorbing/idled. As shown here, the VOC-laden gas enters vessel #1 through valve A, passes through the carbon bed (shown by the shading) and exits through valve B, from whence it passes to the stack. Meanwhile, vessel #2 is in the *desorption* cycle. Steam enters through valve C, flows

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[1] Although steam is the most commonly used regenerant, there are situations where it should *not* be used. An example would be a degreasing operation that emits halogenated VOCs. Steaming might cause the VOCs to decompose .

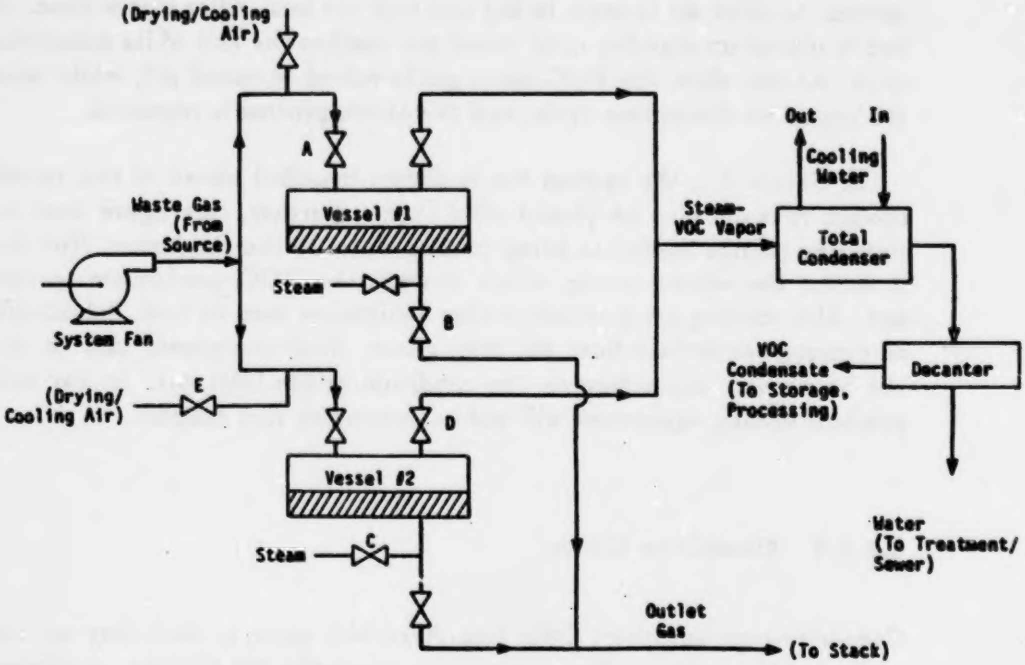

Figure 4.1: Typical Two-Bed, Continuously Operated Fixed-Bed Carbon Adsorber System

through the bed and exits through D. The steam-VOC vapor mixture passes to a condenser, where cooling water condenses the entire mixture. If part of the VOC is immiscible in water, the condensate next passes to a decanter, where the VOC and water layers are separated. The VOC layer is conveyed to storage. If impure, it may receive additional purification by distillation. Depending on its quality (*i.e.*, quantity of dissolved organics), the water layer is usually discharged to a wastewater treatment facility.

Once steaming is completed, valves C and D are closed and valve E is opened, to allow air to enter to dry and cool the bed. After this is done, the bed is placed on standby until vessel #1 reaches the end of its adsorption cycle. At this time, the VOC-laden gas is valved to vessel #2, while vessel #1 begins its desorption cycle, and the above process is repeated.

In Figure 4.1, the system fan is shown installed ahead of the vessels, though it could also be placed after them. Further, this figure does not show the pumps needed to bring cooling water to the condenser. Nor does it depict the solvent pump which conveys the VOC condensate to storage. Also missing are preconditioning equipment used to cool, dehumidify, or remove particulate from the inlet gases. Such equipment may or may not be needed, depending on the condition of the inlet gas. In any case, preconditioning equipment will *not* be covered in this chapter.

4.1.2.2 Cannister Units

Cannister-type adsorbers differ from fixed-bed units, in that they are normally limited to controlling low-volume, (typically 100 ft^3/min, maximum) intermittent gas streams, such as those emitted by storage tank vents, where process economics dictate that either toll regeneration or throw-away cannisters are appropriate. The carbon cannisters are not intended for desorption on-site. However, the carbon may be regenerated at a central facility. Once the carbon reaches a certain VOC content, the unit is shut down, replaced with another, and disposed of or regenerated by the central facility. Each cannister unit consists of a vessel, activated carbon, inlet connection and distributer leading to the carbon bed, and an outlet connection for the purified gas stream.[4] In one design (Calgon's Ventsorb®), 150 lbs of carbon are installed on an 8-inch gravel bed, in a 55-gallon drum. The type of carbon used depends on the nature of the VOC to be treated.

In theory, a cannister unit would remain in service no longer than a regenerable unit would stay in its adsorption cycle. Doing so would help to insure the allowable outlet concentration from being exceeded. In reality, however, poor operating practice may result in the cannister remaining connected until the carbon is near or at saturation. This is because: (1) the carbon (and often the vessel) will probably be disposed of, so there is the temptation to operate it until the carbon is saturated; and (2) unlike fixed-bed units, whose outlet VOC concentrations are usually monitored continuously (via flame ionization detectors, typically), cannisters are usually *not* monitored. Thus, the user can only guess at the outlet loading, and could tend to leave a unit in place longer.

4.1.3 Adsorption Theory

At equilibrium, the quantity of gas that is adsorbed on activated carbon is a function of the adsorption temperature and pressure, the chemical species being adsorbed, and the carbon characteristics, such as carbon particle size and pore structure. For a given adsorbent-VOC combination at a given temperature, an *adsorption isotherm* can be constructed which relates the mass of adsorbate per unit weight of adsorbent ("equilibrium adsorptivity") to the partial pressure of the VOC in the gas stream. The adsorptivity increases with increasing VOC partial pressure and decreases with increasing temperature.

A family of adsorption isotherms having the shape typical of adsorption on activated carbon is plotted in Figure 4.2. This and other isotherms whose shapes are convex upward throughout, are designated "Type I" isotherms. The Freundlich isotherm, which can be fit to a portion of a Type I curve, is commonly used in industrial design.[2]

$$w_e = k\,P^m \tag{4.1}$$

where

w_e = equilibrium adsorptivity (lb adsorbate/lb adsorbent)
P = partial pressure of VOC in gas stream (psia)
k, m = empirical parameters

The treatment of adsorption from gas *mixtures* is complex and beyond

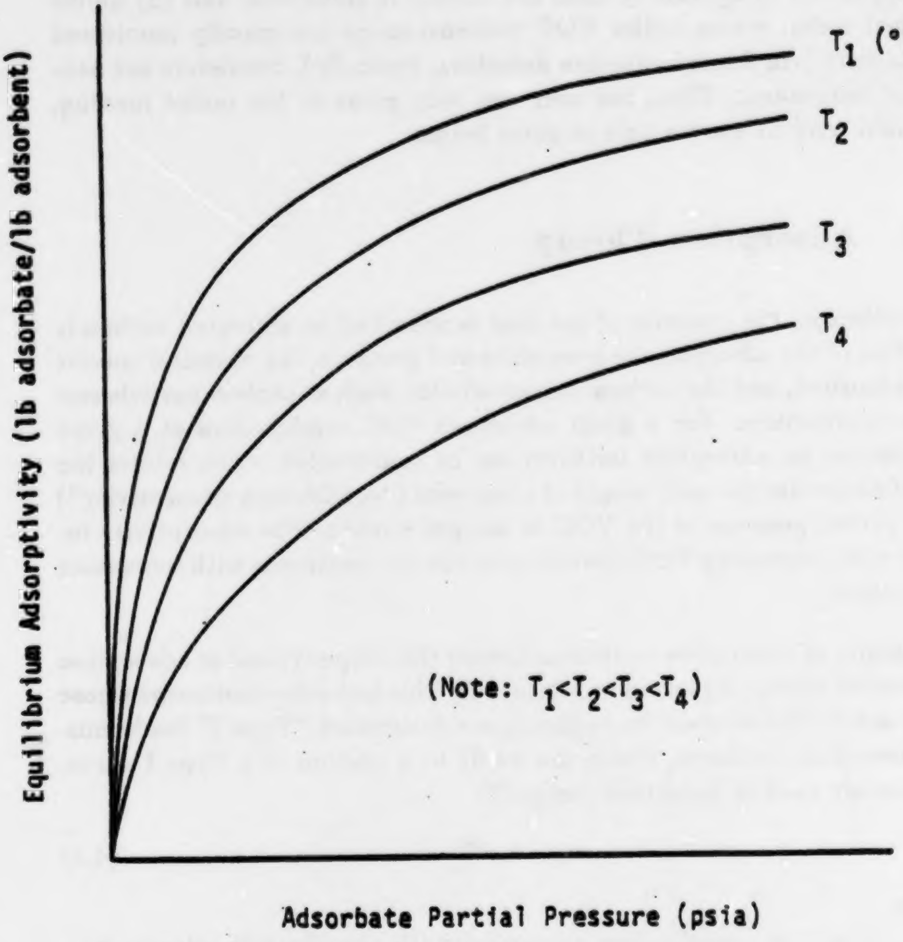

Figure 4.2: Type I Adsorption Isotherms For Hypothetical Adsorbate

the scope of this chapter. Except where the VOC in these mixtures have nearly identical adsorption isotherms, one VOC in a mixture will tend to displace another on the carbon surface. Generally, VOCs with lower vapor pressures will displace those with higher vapor pressures, resulting in the former displacing the latter previously adsorbed. Thus, during the course of the adsorption cycle the carbon's capacity for a higher vapor pressure constituent decreases. This phenomenon should be considered when sizing the adsorber. To be conservative, one would normally base the adsorption cycle requirements on the *least* adsorbable component in a mixture and the desorption cycle on the *most* adsorbable component.[1]

The *equilibrium* adsorptivity is the *maximum amount* of adsorbate the carbon can hold at a given temperature and VOC partial pressure. In actual control systems, however, the entire carbon bed is never allowed to reach equilibrium. Instead, once the outlet concentration reaches a preset limit (the "breakthrough concentration"), the adsorber is shut down for desorption or (in the case of cannister units) replacement and disposal. At the point where the vessel is shut down, the average bed VOC concentration may only be 50% or less of the equilibrium concentration. That is, the carbon bed may be at equilibrium ("saturated") at the gas inlet, but contain only a small quantity of VOC near the outlet.

As Equation 4.1 indicates, the Freundlich isotherm is a power function that plots as a straight line on log-log paper. Conveniently, for the concentrations/partial pressures normally encountered in carbon adsorber operation, most VOC-activated carbon adsorption conforms to Equation 4.1. At very low concentrations, typical of breakthrough concentrations, a linear approximation (on arithmetic coordinates) to the Freundlich isotherm is adequate. *However, the Freundlich isotherm does not accurately represent the isotherm at high gas concentrations and thus should be used with care as such concentrations are approached.*

Adsorptivity data for selected VOCs were obtained from Calgon Corporation, a vendor of activated carbon.[5] The vendor presents adsorptivity data in two forms: a set of graphs displaying equilibrium isotherms [5] and as a modification of the Dubinin-Radushkevich (D-R) equation, a semi-empirical equation that predicts the adsorptivity of a compound based on its adsorption potential and polarizability.[6] In this *Manual*, the modified D-R equation is referred to as the Calgon fifth-order polynomial. The data displayed in the Calgon graphs [5] has been fit to the Freundlich equa-

Table 4.1: Parameters for Selected Adsorption Isotherms[*][a]

		Isotherm Parameters		Range of isotherm[b]
Adsorbate	Adsorption Temp. (°F)	k	m	(psia)
(1) Benzene	77	0.597	0.176	0.0001–0.05
(2) Chlorobenzene	77	1.05	0.188	0.0001–0.01
(3) Cyclohexane	100	0.508	0.210	0.0001–0.05
(4) Dichloroethane	77	0.976	0.281	0.0001–0.04
(5) Phenol	104	0.855	0.153	0.0001–0.03
(6) Trichloroethane	77	1.06	0.161	0.0001–0.04
(7) Vinyl Chloride	100	0.200	0.477	0.0001–0.05
(8) m-Xylene	77	0.708	0.113	0.0001–0.001
	77	0.527	0.0703	0.001–0.05
(9) Acrylonitrile	100	0.935	0.424	0.0001–0.015
(10) Acetone	100	0.412	0.389	0.0001–0.05
(11) Toluene	77	0.551	0.110	0.0001–0.05

[*] Reference [5].

[a] Each isotherm is of the form: $w_e = kP^m$. (See text for definition of terms.) Data are for adsorption on Calgon type "BPL" carbon.

[b] Equations should *not* be extrapolated outside these ranges.

tion. The resulting Freundlich parameters are shown in Table 4.1 for a limited number of chemicals. The adsorbates listed include aromatics (*e.g.*, benzene, toluene), chlorinated aliphatics (dichloroethane), and one ketone (acetone). However, the list is far from all-inclusive.

Notice that a range of partial pressures is listed with each set of parameters, k and m. (Note: In one case (m-xylene) the isotherm was so curvilinear that it had to be split into two parts, each with a different set of parameters.) This is the range to which the parameters apply. Extrapolation beyond this range—especially at the high end—can introduce inaccuracy to the calculated adsorptivity.

But high-end extrapolation may not be necessary, as the following will show. In most air pollution control applications, the system pressure is approximately one atmosphere (14.696 psia). The upper end of the partial pressure ranges in Table 4.1 goes from 0.04 to 0.05 psia. According to Dalton's Law, at a total system pressure of one atmosphere this corresponds to an adsorbate concentration in the waste gas of 2,720 to 3,400 ppmv. Now, as discussed in Section 4.1.2, the adsorbate concentration is usually kept at 25% of the lower explosive limit (LEL).[2] For many VOCs, the LEL ranges from 1 to 1.5 volume %, so that 25% of the LEL would be 0.25 to 0.375% or 2,500 to 3,750 ppmv, which approximates the high end of the partial pressure ranges in Table 4.1.

Finally, each set of parameters applies to a fixed adsorption temperature, ranging from 77° to 104°F. These temperatures reflect typical operating conditions, although adsorption can take place as low as 32°F and even higher than 104°F. As the adsorption temperature increases to much higher levels, however, the equilibrium adsorptivity decreases to such an extent that VOC recovery by carbon adsorption may become economically impractical.

The Calgon fifth-order polynomial is somewhat more accurate than the Freundlich parameters from Table 4.1. The polynomial contains a temperature parameter, and it allows one to estimate adsorption isotherms for compounds not shown in Table 4.1 if pure component data are available. The pure component data required are the saturation pressure, liquid molar volume, and the refractive index. It is, however, somewhat more complex to use than the Freundlich equation. The Calgon fifth-order polynomial is as follows:

The mass loading, w_e, is calculated from

$$w_e = \frac{0.01\,G}{V_m} \times \text{(Molecular Wt of Adsorbate)} \tag{4.2}$$

where

w_e = mass loading, *i.e.*, equilibrium adsorptivity (lb adsorbate per lb carbon)

G = carbon loading at equilibrium (ft^3 liquid adsorbate per 100 lb carbon)

V_m = liquid molar volume of adsorbate (ft^3 per lb-mole).

---

[2] Although, Factory Mutual Insurance will reportedly permit operation at up to 50% of the LEL, if proper VOC monitoring is used.

The carbon loading, G, is calculated from the Calgon fifth-order polynomial

$$log_{10}(G) = A_0 + A_1Y + A_2Y^2 + A_3Y^3 + A_4Y^4 + A_5Y^5 \qquad (4.3)$$

where

$A_0 = 1.71$
$A_1 = -1.46 \times 10^{-2}$
$A_2 = -1.65 \times 10^{-3}$
$A_3 = -4.11 \times 10^{-4}$
$A_4 = +3.14 \times 10^{-5}$
$A_5 = -6.75 \times 10^{-7}$

and Y is calculated from several equations which follow.

The first step in calculating Y is to calculate χ. This can be done by calculating the adsorption potential, ϵ:

$$\epsilon = 0.556\,RT\,\ln(P_s/P_i) \qquad (4.4)$$

where

R = 0.730 (atm-ft^3 per lb-mole-°R)
T = absolute temperature (°R)
P_s = vapor pressure of adsorbate at the temperature T (psia)
P_i = partial pressure of adsorbate (psia).

The χ is calculated from:

$$\chi = \epsilon/(2.303RV_m)$$

By substituting for ϵ in the above equation, χ can alternatively be calculated from:

$$\chi = 0.00890\,(T/V_m)\,\log_{10}(P_s/P_i).$$

The next step in calculating Y is to calculate the relative polarizability, Γ.

$$\Gamma = \Theta_i/\Theta_o$$

where

Θ_i = polarizability of component i per unit volume, where component i is the adsorbate

Θ_o = polarizability of component o per unit volume, where component o is the reference component, n-heptane.

For the adsorbate or the reference compound, using the appropriate refractive index of adsorbate, n, the polarizability is calculated from:

$$\Theta = \frac{n^2 - 1}{n^2 + 2}$$

Once χ and Γ are known, Y can be calculated from:

$$Y = \frac{\chi}{\Gamma} \tag{4.5}$$

Calgon also has a proprietary seventh-order form in which two additional coefficients are added to the Calgon fifth-order polynomial, but the degree of fit reportedly is improved only modestly.[6] Additional sources of isotherm data include the activated carbon vendors, handbooks (such as *Perry's Chemical Engineer's Handbook*), and the literature.

4.2 Design Procedure

4.2.1 Sizing Parameters

Data received from adsorber vendors indicate that the size and purchase cost of a fixed-bed or cannister carbon adsorber system primarily depend on four parameters:

1. The *volumetric flow* of the VOC laden gas passing through the carbon bed(s);
2. The *inlet* and *outlet VOC mass loadings* of the gas stream;
3. The *adsorption time* (*i.e.*, the time a carbon bed remains on-line to adsorb VOC before being taken off-line for desorption of the bed);

4. The *working capacity* of the activated carbon.

In addition, the cost could also be affected by other stream conditions, such as the presence/absence of excessive amounts of particulate, moisture, or other substances which would require the use of extensive pretreatment and/or corrosive-resistant construction materials.

The purchased cost depends to a large extent on the volumetric flow (usually measured in actual ft^3/min). The flow, in turn, determines the size of the vessels housing the carbon, the capacities of the fan and motor needed to convey the waste gas through the system, and the diameter of the internal ducting.

Also important are the VOC inlet and outlet gas stream loadings, the adsorption time, and the working capacity of the carbon. These variables determine the amount and cost of carbon charged to the system initially and, in turn, the cost of replacing that carbon after it is exhausted (typically, five years after startup). Moreover, the amount of the carbon charge affects the size and cost of the auxiliary equipment (condenser, decanter, bed drying/cooling fan), because the sizes of these items are tied to the amount of VOC removed by the bed. The amount of carbon also has a bearing on the size and cost of the vessels.

A carbon adsorber vendor[7] supplied data that illustrate the dependency of the equipment cost on the amount of the carbon charge. Costs were obtained for fixed-bed adsorbers sized to handle three gas flow rates ranging from 4,000 to 100,000 scfm and to treat inlet VOC (toluene) concentrations of 500 and 5,000 ppm. Each adsorber was assumed to have an eight hour adsorption time. As one might expect, the equipment costs for units handling higher gas flow rates were higher than those handling lower gas flow rates. Likewise, at each of the gas flow rates, the units sized to treat the 5,000 ppm VOC streams had higher equipment costs than those sized to treat the 500 ppm concentration. These cost differences ranged from 23 to 29% and averaged 27%. These higher costs were partly needed to pay for the additional carbon required to treat the higher concentration streams. But some of these higher costs were also needed for enlarging the adsorber vessels to accommodate the additional carbon and for the added structural steel to support the larger vessels. Also, larger condensers, decanters, cooling water pumps, etc., were necessary to treat the more concentrated streams. (See Section 4.3.)

The VOC inlet loading is set by the source parameters, while the *outlet* loading is set by the VOC emission limit. (For example, in many states, the average VOC outlet concentration from adsorbers may not exceed 25 ppm.)

4.2.2 Determining Adsorption and Desorption Times

The relative times for adsorption and desorption and the adsorber bed configuration (*i.e.*, whether single or multiple and series or parallel adsorption beds are used) establish the adsorption/desorption cycle profile. The cycle profile is important in determining carbon and vessel requirements and in establishing desorption auxiliary equipment and utility requirements. An example will illustrate. In the simplest case, an adsorber would be controlling a process which emits a relatively small amount of VOC intermittently —say, during one 8-hour shift per day. During the remaining 16 hours the system would either be desorbing or on standby. Properly sized, such a system would only require a single bed, which would contain enough carbon to treat eight hours worth of gas flow at the specified inlet concentration, temperature, and pressure. Multiple beds, operating in parallel, would be needed to treat large gas flows (>100,000 actual ft^3/min, generally)[7], as there are practical limits to the sizes to which adsorber vessels can be built. But, regardless of whether a single bed or multiple beds were used, the system would only be on-line for part of the day.

However, if the process were operating *continuously* (24 hours), an extra carbon bed would have to be installed to provide adsorptive capacity during the time the first bed is being regenerated. The amount of this extra capacity must depend on the number of carbon beds that would be adsorbing at any one time, the length of the adsorption period relative to the desorption period, and whether the beds were operating in parallel or in series. If one bed were adsorbing, a second would be needed to come on-line when the first was shut down for desorption. In this case, 100% extra capacity would be needed. Similarly, if *five* beds in parallel were operating in a staggered adsorption cycle, only *one* extra bed would be needed and the extra capacity would be 20% (*i.e.*, 1/5)—provided, of course, that the adsorption time were at least five times as long as the desorption time. The relationship between adsorption time, desorption time, and the required extra capacity can be generalized.

$$M_c = M_{c_I} \times f \tag{4.6}$$

where

M_c, M_{c_I} = amounts of carbon required for *continuous* or *intermittent* control of a given source, respectively (lbs)

f = extra capacity factor (dimensionless)

This equation shows the relationship between M_c and M_{c_I}. Section 4.2.3 shows how to calculate these quantities.

The factor, f, is related to the number of beds adsorbing (N_A) and desorbing (N_D) in a *continuous* system as follows:

$$f = 1 + \frac{N_D}{N_A} \tag{4.7}$$

(Note: N_A is also the number of beds in an intermittent system that would be adsorbing at any given time. The *total* number of beds in the system would be $N_A + N_D$.)

It can be shown that the number of desorbing beds required in a continuous system (N_D) is related to the desorption time (θ_D), adsorption time (θ_A), and the number of adsorbing beds, as follows:

$$\theta_D \le \theta_A \left(\frac{N_D}{N_A}\right) \tag{4.8}$$

(Note: θ_D is the total time needed for bed regeneration, drying, and cooling.)

For instance, for an eight hour adsorption time, in a continuously operated system of seven beds (six adsorbing, one desorbing) θ_D would have to be $1\frac{1}{3}$ hours or less (8 hours/6 beds). Otherwise, additional beds would have to be added to provide sufficient extra capacity during desorption.

4.2.3 Estimating Carbon Requirement

4.2.3.1 Overview of Carbon Estimation Procedures

Obtaining the carbon requirement (M_c or M_{c_T}) is not as straightforward as determining the other adsorber design parameters. When estimating the carbon charge, the sophistication of the approach used depends on the data and calculational tools available.

One approach for obtaining the carbon requirement is a rigorous one which considers the unsteady-state energy and mass transfer phenomena occurring in the adsorbent bed. Such a procedure necessarily involves a number of assumptions in formulating and solving the problem. Such a procedure is beyond the scope of this *Manual* at the present time, although ongoing work in the Agency is addressing this approach.

In preparing this chapter of the *Manual*, we have adopted a rule-of-thumb procedure for estimating the carbon requirement. This procedure, while approximate in nature, appears to have the acceptance of vendors and field personnel. It is sometimes employed by adsorber vendors to make rough estimates of carbon requirement and is relatively simple and easy to use. It normally yields results incorporating a safety margin, the size of which depends on the bed depth (short beds would have less of a safety margin than deep beds), the effectiveness of regeneration, the particular adsorbate and the presence or absence of impurities in the stream being treated.

4.2.3.2 Carbon Estimation Procedure Used in *Manual*

The rule-of-thumb carbon estimation procedure is based on the "working capacity" (w_c, lb VOC/lb carbon). This is the *difference* per unit mass of carbon between the amount of VOC on the carbon at the end of the adsorption cycle and the amount remaining on the carbon at the end of the desorption cycle. It should *not* be confused with the "equilibrium capacity" (w_e) defined above in section 4.1.3. Recall that the equilibrium capacity measures the capacity of virgin activated carbon when the VOC has been in contact with it (at a constant temperature and partial pressure) long enough to reach equilibrium. In adsorber design, it would not be feasible to allow the bed to reach equilibrium. If it were, the outlet concentration would rapidly increase beyond the allowable outlet (or "breakthrough") concentration until the outlet concentration reached the inlet concentration. During this period the adsorber would be violating the emission limit.

The working capacity is some fraction of the equilibrium capacity. Like the equilibrium adsorptivity, the working capacity depends upon the temperature, the VOC partial pressure, and the VOC composition. The working capacity also depends on the flow rate and the carbon bed parameters.

The working capacity, along with the adsorption time and VOC inlet loading, is used to compute the carbon requirement for a cannister adsorber or for an *intermittently* operated fixed-bed adsorber as follows:

$$M_{c_I} = \frac{m_{voc}}{w_c}\theta_A \qquad (4.9)$$

where m_{voc} = VOC inlet loading (lb/h)

Combining this with Equations 4.6 and 4.7 yields the general equation for estimating the system total carbon charge for a *continuously* operated system:

$$M_c = \frac{m_{voc}}{w_c}\theta_A\left(1 + \frac{N_D}{N_A}\right) \qquad (4.10)$$

Values for w_c may be obtained from knowledge of operating units. If no value for w_c is available for the VOC (or VOC mixture) in question, the working capacity may be estimated at 50% of the equilibrium capacity, as follows:

$$w_c \approx 0.5 w_{e(max)} \qquad (4.11)$$

where $w_{e(max)}$ = the equilibrium capacity (lb VOC/lb carbon) taken at the adsorber inlet (*i.e.*, the point of maximum VOC concentration).

(Note: To be conservative, this 50% figure should be lowered if short desorption cycles, very high vapor pressure constituents, high moisture contents, significant amounts of impurities, or difficult-to-desorb VOCs are involved. Furthermore, the presence of strongly adsorbed impurities in the inlet VOC stream may significantly shorten carbon life.)

As Equation 4.10 shows, the carbon requirement is directly proportional to the adsorption time. This would tend to indicate that a system could be designed with a shorter adsorption time to minimize the carbon requirement (and equipment cost). There is a trade-off here not readily apparent from Equation 4.10, however. Certainly, a shorter adsorption time would require less carbon. But, it would also mean that a carbon bed would have to be desorbed more frequently. This would mean that the regeneration steam would have to be supplied to the bed(s) more frequently to remove (in the long run) the same amount of VOC. Further, each time the bed is regenerated the steam supplied must heat the vessel and carbon, as well as drive off the adsorbed VOC. And the bed must be dried and cooled after each desorption, regardless of the amount of VOC removed. Thus, if the bed is regenerated too frequently, the bed drying/cooling fan must operate more often, increasing its power consumption. Also, more frequent regeneration tends to shorten the carbon life. As a rule-of-thumb, the optimum regeneration frequency for fixed-bed adsorbers treating streams with moderate to high VOC inlet loadings is once every 8 to 12 hours.[1]

4.3 Estimating Total Capital Investment

Entirely different procedures should be used to estimate the purchased costs of fixed-bed and cannister-type adsorbers. Therefore, they will be discussed separately.

4.3.1 Fixed-Bed Systems

As indicated in the previous section, the purchased cost is a function of the volumetric flow rate, VOC inlet and outlet loadings, the adsorption time, and the working capacity of the activated carbon. As Figure 4.1 shows, the adsorber system is made up of several different items. Of these, the adsorber vessels and the carbon comprise from one-half to nearly 90% of the total equipment cost. (See Section 4.3.1.3.) There is also auxiliary equipment, such as fans, pumps, condensers, decanters, and internal piping. But because these usually comprise a small part of the total purchased cost, they may be "factored" from the costs of the carbon and vessels without introducing significant error. The costs of these major items will be considered separately.

4.3.1.1 Carbon Cost

This cost (C_c,\$) is simply the product of the initial carbon requirement (M_c) and the current price of carbon. As adsorber vendors buy carbon in very large quantities (million-pound lots or larger), their cost is somewhat lower than the list price. A typical vendor cost is \$2.00/lb (fall 1989 dollars).[8] Thus:

$$C_c = 2.00 M_c \tag{4.12}$$

4.3.1.2 Vessel Cost

The cost of an adsorber vessel is primarily determined by its dimensions which, in turn, depend upon the amount of carbon it must hold and the superficial gas velocity through the bed that must be maintained for optimum adsorption. The desired superficial velocity is used to calculate the cross-sectional area of the bed perpendicular to the gas flow. An acceptable superficial velocity is established empirically, considering desired removal efficiency, the carbon particle size and bed porosity, and other factors. For example, one adsorber vendor recommends a superficial bed velocity of 85 ft/min[7], while an activated carbon manufacturer cautions against exceeding 60 ft/min in systems operating at one atmosphere.[5] Another vendor uses a 65 ft/min superficial face velocity in sizing its adsorber vessels.[8] Lastly, there are practical limits to vessel dimensions which also influence their sizing. That is, due to shipping restrictions, vessel diameters rarely exceed 12 feet, while their length is generally limited to 50 feet.[8]

The cost of a vessel is usually correlated with its weight. However, as the weight is often difficult to obtain or calculate, the cost may be estimated from the external surface area. This is true because the vessel material cost—and the cost of fabricating that material—is directly proportional to its surface area. The surface area (S, ft^2) of a vessel is a function of its length (L, ft) and diameter (D, ft), which in turn, depend upon the superficial bed face velocity, the L/D ratio, and other factors.

Most commonly, adsorber vessels are cylindrical in shape and erected horizontally (as in Figure 4.1). Vessels configured in this manner are generally subjected to the constraint that the carbon volume occupies no more than 1/3 of the vessel volume [7,8]. It can be shown that this constraint limits the bed depth to no more than

$$\text{Maximum bed depth} \approx \frac{\pi D}{12}. \tag{4.13}$$

The vessel length, L, and diameter, D, can be estimated by solving two relationships, namely, (1) the equation relating carbon volume, and thus vessel volume, to L and D, and (2) the equation relating volumetric flow rate, superficial velocity, and cross-section normal to flow. If one assumes that the carbon bulk density is 30 lb/ft^3, then one can show that:

$$D = \frac{0.127 M_c' v_b}{Q'} \tag{4.14}$$

$$\text{and:} \quad L = \frac{7.87}{M_c'}\left(\frac{Q'}{v_b}\right)^2 \tag{4.15}$$

where

D = vessel diameter (ft)
L = vessel length (ft)
v_b = bed superficial velocity (ft/min)
M_c' = carbon requirement *per vessel* (lbs)
Q' = volumetric flow rate *per adsorbing vessel* (acfm)

Because the constants in equations 4.14 and 4.15 are not dimensionless, one must be careful to use the units specified in these equations.

Although other design considerations can result in different values of L and D, these equations result in L and D which are acceptable from the standpoint of "study" cost estimation for horizontal, cylindrical vessels which are larger than 2-3 feet in diameter.

The carbon requirement and flow rate for each adsorber vessel can be calculated as follows:

$$M_c' = \frac{M_c}{(N_A + N_D)}$$

$$Q' = \frac{Q}{N_A}.$$

At gas flow rates (Q') of less than 9,000 scfm, it is usually more feasible to erect the adsorber vessels vertically instead of horizontally.[8] If so, the vessel diameter can be calculated from the volumetric flow rate per adsorbing vessel and the bed superficial velocity as follows:

$$D = \left(\frac{4Q'}{\pi v_b}\right)^{1/2} \tag{4.16}$$

The vertical vessel length will depend principally on the carbon bed thickness. Additional space must be included below the carbon bed for bed support and above and below the bed for distribution and disengaging of the gas stream and for physical access to the carbon bed. In smaller diameter vessels, access to both sides of the bed is usually not required. However, 1 to $1\frac{1}{2}$ feet must be provided on *each* side for gas distribution and disengagement, or 2 to 3 feet overall. For longer vessels, 2 to 3 feet at each end of the vessel is typically provided for access space.

Given the mass of carbon in the bed, the carbon bulk density, and the bed diameter (i.e., the cross-sectional area normal to flow), determining the carbon bed thickness is straightforward using the following equation:

$$t_b = \frac{\text{volume of carbon}}{\text{cross-sectional area normal to flow}} = \frac{M_c'/\rho_b}{Q'/v_b} \tag{4.17}$$

where

ρ = carbon bulk density (lb/ft^3, assume 30 lb/ft^3)

The vessel length is, therefore,

$$L = t_b + t_{a,g} \tag{4.18}$$

where

$t_{a,g}$ = access/ gas distribution allowance
= 2 to 6 feet (depending on vertical vessel diameter)

Finally, use the following equation to calculate the surface area of either a horizontal or vertical vessel:

$$S = \pi D(L + D/2) \tag{4.19}$$

Similar equations can be developed for other vessel shapes, configurations, etc.

Based on vendor data, we developed a correlation between adsorber vessel cost and surface area:[8]

$$C_v = 271\, S^{0.778} \tag{4.20}$$

where C_v = vessel cost (fall 1989 \$), F.O.B. vendor
and $97 \leq S \leq 2,110$ ft^2.

These units would be made of 304 stainless steel, which is the most common material used in fabricating adsorber vessels.[7,8] However, to obtain the cost of a vessel fabricated of another material, multiply C_v by an adjustment factor (F_m). A few of these factors are listed below:

Material	F_m Factor	Reference(s)
Stainless steel, 316	1.3	[7,8,9]
Carpenter 20 CB-3	1.9	[9]
Monel-400	2.3	[7,9]
Nickel-200	3.2	[9]
Titanium	4.5	[9]

4.3.1.3 Total Purchased Cost

As stated earlier, the costs of such items as the fans, pumps, condenser, decanter, instrumentation, and internal piping can be factored from the sum of the costs for the carbon and vessels. Based on four data points derived from costs supplied by an equipment vendor [8], we found that, depending on the total gas flow rate (Q), the ratio (R_c) of the total adsorber equipment cost to the cost of the vessels and carbon ranged from 1.14 to 2.24. These data points spanned a gas flow rate range of approximately 4,000 to 500,000 acfm. The following regression formula fit these four points:

$$R_c = 5.82\, Q^{-0.133} \tag{4.21}$$

where

$4,000 \leq Q \text{ (acfm)} \leq 500,000$
Correlation coefficient (r) = 0.872

The total adsorber equipment cost (C_A) would be the product of R_c and the sum of the carbon and vessel costs, or:

$$C_A = R_c\,[C_c + (N_A + N_D)C_v] \tag{4.22}$$

4.3.1.4 Total Capital Investment

As discussed in Chapter 2, in the methodology used in this *Manual*, the total capital investment (TCI) is estimated from the total purchased cost via an overall direct/indirect installation cost factor. A breakdown of that factor for carbon adsorbers is shown in Table 4.2. As Chapter 2 indicates, the TCI also includes costs for land, working capital, and off-site facilities, which are not included in the direct/indirect installation factor. However, as these items are rarely required with adsorber systems, they will not be considered here. Further, no factors have been provided for site preparation (SP) and buildings (Bldg.), as these site-specific costs depend very little on the purchased equipment cost.

Note that the installation factor is applied to the total purchased equipment cost, which includes the cost of such auxiliary equipment as the stack and external ductwork and such costs as freight and sales taxes (if applicable). ("External ductwork" is that ducting needed to convey the exhaust gas from the source to the adsorber system, and then from the adsorber to the stack. Costs for ductwork and stacks are shown elsewhere in this *Manual.*) Normally, the adjustment would also cover the instrumentation cost, but this cost is usually included with the adsorber equipment cost. Finally, note that these factors reflect "average" installation conditions and could vary considerably, depending upon the installation circumstances.

4.3.2 Cannister Systems

Once the carbon requirement is estimated using the above procedure, the number of cannisters is determined. This is done simply by dividing the total carbon requirement (M_c) by the amount of carbon contained by each cannister (typically, 150 lbs.). This quotient, rounded to the next *highest* digit, yields the required number of cannisters to control the vent in question.

Costs for a typical cannister (Calgon's Ventsorb®) are listed in Table 4.3. These costs include the vessel, carbon, and connections, but do not include taxes, freight, or installation charges. Note that the cost per unit decreases as the quantity purchased increases. Each cannister contains Calgon's "BPL" carbon (4 × 10 mesh), which is commonly used in indus-

Table 4.2: Capital Cost Factors for Carbon Adsorbers[a]

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
Adsorber + auxiliary equipment[b]	As estimated, A
Instrumentation[c]	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.08 B
Handling & erection	0.14 B
Electrical	0.04 B
Piping	0.02 B
Insulation	0.01 B
Painting	0.01 B
Direct installation costs	0.30 B
Site preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Costs, DC	1.30 B + SP + Bldg.
Indirect Costs (installation)	
Engineering	0.10 B
Construction and field expenses	0.05 B
Contractor fees	0.10 B
Start-up	0.02 B
Performance test	0.01 B
Contingencies	0.03 B
Total Indirect Costs, IC	0.31 B
Total Capital Investment = DC + IC	1.61 B + SP + Bldg.

[a] Reference [10].

[b] Ductwork and any other equipment normally not included with unit furnished by adsorber vendor.

[c] Instrumentation and controls often furnished with the adsorber, and thus included in the EC.

Table 4.3: Equipment Costs (Spring 1986 $) for a Typical Cannister Adsorber[a]

Quantity	Equipment Cost (each)[b]
1-3	$687
4-9	659
10-29	622
≥30	579

[a] Reference [4].

[b] These costs are F.O.B., Pittsburgh, PA. They do *not* include taxes and freight charges.

trial adsorption. However, to treat certain VOCs, more expensive speciality carbons (*e.g.*, "FCA 4 × 10") are needed. These carbons can increase the equipment cost by 60% or more.[4] As is indicated in the caption of Table 4.3, these prices are in Spring 1986 dollars. Since then, however, the prices of these cannisters have increased modestly—approximately 10%.[11]

As fewer installation materials and labor are required to install a cannister unit than a fixed-bed system, the composite installation factor is consequently lower. The only costs required are those needed to place the cannisters at, and connect them to, the source. This involves a small amount of piping only; little or no electrical work, painting, foundations, or the like would be needed. *Twenty percent* of the sum of the cannister(s) cost, freight charges, and applicable sales taxes would cover this installation cost.

4.4 Estimating Total Annual Cost

As Chapter 2 of this *Manual* explains, the total annual cost is comprised of three components: *direct costs*, *indirect costs*, and *recovery credits*. These will be considered separately.

4.4.1 Direct Annual Costs

These include the following expenditures: steam, cooling water, electricity, carbon replacement, operating and supervisory labor, and maintenance labor and materials. Of these, only ***electricity*** and ***solid waste disposal*** would apply to the cannister-type adsorbers.

4.4.1.1 Steam

As explained in section 4.1, steam is used during the desorption cycle. The quantity of steam required will depend on the amount of carbon in the vessel, the vessel dimensions, the type and amount of VOC adsorbed, and other variables. Experience has shown that the steam requirement ranges from approximately 3 to 4 lbs of steam/lb of adsorbed VOC.[7,8] Using the midpoint of this range, we can develop the following expression for the annual steam cost:

$$C_s = 3.50 \times 10^{-3} m_{voc} \theta_s \, p_s \qquad (4.23)$$

where

C_s = steam cost (\$/yr)
θ_s = system operating hours (h/yr)
m_{voc} = VOC inlet loading (lbs/h)
p_s = steam price (\$/thous. lbs)

If steam price data are unavailable, one can estimate its cost at 120% of the fuel cost. For example, if the local price of natural gas were \$5.00/million BTU, the estimated steam price would be \$6.00/million BTU which is approximately \$6.00/thousand lbs. (The 20% factor covers the capital and annual costs of producing the steam.)

4.4.1.2 Cooling Water

Cooling water is consumed by the condenser in which the steam-VOC mixture leaving the desorbed carbon bed is totally condensed. Most of the condenser duty is comprised of the latent heat of vaporization (ΔH_v) of the steam and VOC. As the VOC ΔH_v are usually small compared to the steam ΔH_v (about 1000 BTU/lb), the VOC ΔH_v may be ignored. So may the sensible heat of cooling the water-VOC condensate from the condenser inlet temperature (about 212°F) to the outlet temperature. Therefore, the cooling water requirement is essentially a function of the steam usage and the allowable temperature rise in the coolant, which is typically 30° to 40°F.[7] Using the average temperature rise (35°F), we can write:

$$C_{cw} = 3.43\frac{C_s}{p_s}p_{cw} \tag{4.24}$$

where

C_{cw} = cooling water cost ($/yr)
p_{cw} = cooling water price ($/thous. gal.)

If the cooling water price is unavailable, use $0.15 to $0.30/thousand gallons.

4.4.1.3 Electricity

In fixed-bed adsorbers, electricity is consumed by the system fan, bed drying/cooling fan, cooling water pump, and solvent pump(s). Both the system and bed fans must be sized to overcome the pressure drop through the carbon beds. But, while the system fan must continuously convey the total gas flow through the system, the bed cooling fan is only used during a part of the desorption cycle (one-half hour or less).

For both fans, the horsepower needed depends both on the gas flow and the pressure drop through the carbon bed. The pressure drop through the bed (ΔP_b) depends on several variables, such as the adsorption temperature, bed velocity, bed characteristics (*e.g.*, void fraction), and thickness. But, for a given temperature and carbon, the pressure drop *per unit thickness* depends solely on the gas velocity. For instance, for Calgon's "PCB" carbon (4 × 10 mesh), the following relationship holds:[5]

$$\Delta P_b/t_b = 0.03679v_b + 1.107 \times 10^{-4}v_b^2 \tag{4.25}$$

where

$\Delta P_b/t_b$ = pressure drop through bed (inches of water/foot of carbon)
v_b = superficial bed velocity (ft/min)

As Equation 4.17 shows, the bed thickness (t_b, ft) is the quotient of the bed volume (V_b) and the bed cross-sectional area (A_b). For a 30 lb/ft^3 carbon bed density, this becomes:

$$t_b = \frac{V_b}{A_b} = \frac{0.0333M_c'}{A_b} \qquad (4.26)$$

(For vertically erected vessels, $A_b = Q'/v_b$, while for horizontally erected cylindrical vessels, $A \approx LD$.) Once ΔP_b is known, the system fan horsepower requirement (hp_{sf}) can be calculated:

$$hp_{sf} = 2.50 \times 10^{-4}\, Q\, \Delta P_s \qquad (4.27)$$

where

Q = gas volumetric flow through system (acfm)
ΔP_s = total system pressure drop = $\Delta P_b + 1$

(The extra inch accounts for miscellaneous pressure losses through the external ductwork and other parts of the system.[7] However, if extra long duct runs and/or preconditioning equipment are needed, the miscellaneous losses could be much higher.)

This equation incorporates a fan efficiency of 70% and a motor efficiency of 90%, or 63% overall.

The horsepower requirement for the bed drying/cooling fan (hp_{cf}) is computed similarly. While the bed fan pressure drop would still be ΔP_b, the gas flow and operating times would be different. For typical adsorber operating conditions, the drying/cooling air requirement would be 50 to 150 ft^3/lb carbon, depending on the bed moisture content, required temperature drop, and other factors. The operating time (θ_{cf}) would be the product of the drying/cooling time per desorption cycle and the number of cycles per year. It can be shown that:

$$\theta_{cf} = 0.4\theta_D(N_A\theta_s/\theta_A) \qquad (4.28)$$

(The "0.4" allows for the fact that as a rule-of-thumb, approximately 40% of the desorption cycle is used for bed drying/cooling.)

The cooling water pump horsepower requirement (hp_{cwp}) would be computed as follows:

$$hp_{cwp} = \frac{2.52 \times 10^{-4} q_{cw} H\, s}{\eta} \qquad (4.29)$$

where

q_{cw} = cooling water flow (gal/min)
H = required head (nominally 100 feet of water)
s = specific gravity of fluid relative to water at 60°F
η = combined pump-motor efficiency.

The annual operating hours for the cooling water pump (θ_{cwp}) would be computed using Equation 4.28, after substituting "0.6" for 0.4. The 0.6 factor accounts for the fact that the cooling water pump is only used during the steaming portion of the regeneration cycle, while the condenser is in operation.

Equation 4.29 may also be used to compute the solvent pump horsepower requirement. In the latter case, the flow (q_s) would be different, of course, although the same head—100 ft. of water—could be used. The specific gravity would depend on the composition and temperature of the condensed solvent. For example, the specific gravity of toluene at 100°F would be approximately 0.86 at 70°F. (However, the solvent pump horsepower is usually very small—usually < 0.1 hp.—so its electricity consumption can usually be neglected.)

Once the various horsepowers are calculated, the electricity usage (in kWh) is calculated, by multiplying each horsepower value by 0.746 (the factor for converting hp to kilowatts) and the number of hours each fan or pump operates annually. For the system fan, the hours would be the annual operating hours for the system (θ_s). But, as discussed above, the operating times for the bed drying/cooling fan and cooling water pump would be different.

To obtain the annual electricity cost, simply multiply kWh by the electricity price (in $/kWh) that applies to the facility being controlled.

For cannister units, use equation 4.27 to calculate the fan horsepower requirement. However, instead of ΔP_b, use the following to compute the *total* cannister pressure drop (ΔP_c, inches of water):[4]

$$\Delta P_c = 0.0471 Q_c + 9.29 \times 10^{-4} Q_c^2 \qquad (4.30)$$

where Q_c = flow through the cannister (acfm).

4.4.1.4 Carbon Replacement

As discussed above, the carbon has a different economic life than the rest of the adsorber system. Therefore, its replacement cost must be calculated separately. Employing the procedure detailed in Chapter 2, we have:

$$CRC_c = CRF_c(1.08C_c + C_{cl}) \tag{4.31}$$

where

CRF_c = capital recovery factor for the carbon
1.08 = taxes and freight factor
C_c, C_{cl} = initial cost of carbon (F.O.B. vendor) and carbon replacement labor cost, respectively ($)

The replacement labor cost covers the labor cost for removing spent carbon from vessels and replacing it with virgin or regenerated carbon. The cost would vary with the amount of carbon being replaced, the labor rates, and other factors. For example, to remove and replace a 50,000 pound carbon charge would require about 16 person-days, which, at typical wage rates, is equivalent to approximately $0.05/lb replaced.[12]

A typical life for the carbon is *five* years. However, if the inlet contains VOCs that are very difficult to desorb, tend to polymerize, or react with other constituents, a shorter carbon lifetime—perhaps as low as *two* years—would be likely.[1] For a five-year life and 10% interest rate, $CRF_c = 0.2638$.

4.4.1.5 Solid Waste disposal

Disposal costs are rarely incurred with fixed-bed adsorbers, because the carbon is almost always regenerated in place, not discarded. In certain cases, the carbon in cannister units is also regenerated, either off-site or at a central regeneration facility on-site. However, most cannister adsorbers are disposed of once they become saturated. The entire cannister—carbon, drum, connections, etc.—is shipped to a secure landfill. The cost of landfill disposal could vary considerably, depending on the number of cannisters disposed of, the location of the landfill, etc. Based on data obtained from two large landfills, for instance, the disposal cost would range from approximately $35 to $65 per cannister excluding transportation costs.[13,14]

4.4.1.6 Operating and Supervisory Labor

The operating labor for adsorbers is relatively low, as most systems are automated and require little attention. *One-half* operator hour per shift is typical.[10] The annual labor cost would then be the product of this labor requirement and the operating labor wage rate ($/h) which, naturally, would vary according to the facility location, type of industry, etc. Add to this 15% to cover supervisory labor, as Chapter 2 suggests.

4.4.1.7 Maintenance Labor and Materials

Use 0.5 hours/shift for maintenance labor [10] and the applicable maintenance wage rate. If the latter data are unavailable, estimate the maintenance wage rate at 110% of the operating labor rate, as Chapter 2 suggests. Finally, for maintenance materials, add an amount equal to the maintenance labor, also per Chapter 2.

4.4.2 Indirect Annual Costs

These include such costs as capital recovery, property taxes, insurance, overhead, and administrative costs ("G&A"). The capital recovery cost is based on the equipment lifetime and the annual interest rate employed. (See Chapter 2 for a thorough discussion of the capital recovery cost and the variables that determine it.) For adsorbers, the system lifetime is typically *ten years*, except for the carbon, which, as stated above, typically needs to be replaced after *five* years. Therefore, when figuring the system capital recovery cost, one should base it on the installed capital cost *less the cost of replacing the carbon* (*i.e.*, the carbon cost plus the cost of labor necessary to replace it). Substituting the initial carbon and replacement labor costs from equation 4.31, we obtain:

$$CRC_s = [TCI - (1.08C_c + C_d)]\,CRF_s \qquad (4.32)$$

where

CRC_s = capital recovery cost for adsorber system ($/yr)
TCI = total capital investment ($)
1.08 = taxes and freight factor
C_c, C_d = initial carbon cost (F.O.B. vendor) and carbon replacement cost, respectively ($)
CRF_s = capital recovery factor for adsorber system (defined in Chapter 2).

For a ten-year life and a 10% annual interest rate, the CRF_s would be 0.1628.

As Chapter 2 indicates, the suggested factor to use for property taxes, insurance, and administrative charges is 4% of the TCI. Finally, the overhead is calculated as 60% of the sum of operating, supervisory, and maintenance labor, and maintenance materials.

The above procedure applies to *cannister* units as well, except that, in most cases, the carbon is not replaced—the entire unit is. Cannisters are generally used in specialized applications. The piping and ducting cost can usually be considered a capital investment with a useful life of ten years. However, whether the cannister itself would be treated as a capital or an operating expense would depend on the particular application and would need to be evaluated on a case-by-case basis.

4.4.3 Recovery Credits

These apply to the VOC which is adsorbed, then desorbed, condensed, and separated from the steam condensate. If the recovered VOC is sufficiently pure, it can be sold. However, if the VOC layer contains impurities or is a mixture of compounds, it would require further treatment, such as distillation. Purification and separation costs are beyond the scope of this chapter. Needless to say, the costs of these operations would offset the revenues generated by the sale of the VOC. Finally, as an alternative to reselling it, the VOC could be burned as fuel and valued accordingly. In any case, the following equation can be used to calculate these credits:

$$RC = m_{\mathrm{voc}}\theta_s p_{\mathrm{voc}}E \qquad (4.33)$$

where

RC = recovery credit ($/yr)
m_{voc} = VOC inlet loading (lbs/h)
θ_s = system operating hours (h/yr)
p_{voc} = resale value of the recovered VOC ($/lb)
E = adsorber VOC control efficiency

By definition, the efficiency (E) is the difference between the inlet and outlet VOC mass loadings, divided by the inlet loading. However, during an adsorption cycle the outlet VOC loading will increase from essentially zero at the start of the cycle to the breakthrough concentration at the end of the cycle. Because the efficiency is a function of time, it should be calculated via integration over the length of the adsorption cycle. To do this would require knowledge of the temporal variation of the outlet loading during the adsorption cycle. If this knowledge is not available to the *Manual* user, a *conservative* approximation of the efficiency may be made by setting the outlet loading equal to the breakthrough concentration.

4.4.4 Total Annual Cost

Finally, as explained in Chapter 2, the total annual cost (TAC) is the sum of the direct and indirect annual costs, less any recovery credits, or:

$$TAC = DC + IC - RC \qquad (4.34)$$

4.4.5 Example Problem

A source at a printing plant emitting 100 lb/h of toluene is to be controlled by a carbon adsorber. The plant proposes to operate the adsorber in a continuous mode for 8,640 h/yr (360 days). While operating, two carbon beds will be adsorbing, while a third will be desorbing/on stand by. For its convenience, the plant has selected adsorption and desorption times of 12 and 5 hours, respectively. The total waste gas flow is 10,000 acfm at the adsorber inlet conditions (one atmosphere and 77°F). The waste gas contains negligible quantities of particulate matter and moisture. Further, the applicable VOC regulation requires the adsorber to achieve a mean removal efficiency of 98% during the entire adsorption cycle. Finally, assume that

the recovered toluene is recycled at the source. Estimate the total capital investment and total annual cost for the adsorber system.

Carbon Working Capacity: At the stated flow and pollutant loading, the toluene inlet concentration is 710 ppm. This corresponds to a partial pressure of 0.0104 psia. Substituting this partial pressure and the toluene isotherm parameters (from Table 4.1) into equation 4.1, we obtain an equilibrium capacity of 0.333 lb/lb. By applying the rule-of-thumb discussed above (page 4-19), we obtain a working capacity of 0.167 lb/lb (*i.e.*, 0.333/2).

Carbon Requirement: As stated above, this adsorber would have two beds on-line and a third off-line. Is this a reasonable assumption? Equation 4.8 can answer this question. Substitution of the adsorption time and numbers of adsorbing and desorbing beds yields:

$$\text{Desorption time} = \theta_D \leq \theta_A(N_D/N_A) = 12\,\text{h}\,(1/2) = 6\text{ h}.$$

Because the stated desorption time (5 hours) is less than 6 hours, the proposed bed configuration is feasible. Next, calculate the carbon requirement (M_c) from equation 4.10:

$$M_c = \frac{m_{\text{voc}}}{w_c}\,\theta_A\left(1 + \frac{N_D}{N_A}\right) = \frac{100\,\text{lb/h}\,(12\,\text{h})(1 + 1/2)}{0.167\,\text{lb/lb}} = 10,800\,\text{lbs}.$$

From equation 4.12, the carbon cost is:

$$C_c = 2.00 M_c = \$21,600.$$

Adsorber Vessel Dimensions and Cost: Assume that the vessels will be erected horizontally and select a superficial bed velocity (v_b) of 75 ft/min. Next, calculate the vessel diameter (D), length (L), and surface area (S) from equations 4.14, 4.15, and 4.19, respectively. [Note: In these equations, $M_c' = M_c/(N_A + N_D) = 3,600$ lb and $Q' = Q/N_A = 5,000$ acfm.]

$$D = \frac{0.127 M_c' v_b}{Q'} = \frac{0.127(3,600)(75)}{5,000} = 6.86\text{ ft}$$

$$L = \frac{7.87}{M_c'}\left(\frac{Q'}{v_b}\right)^2 = \frac{7.87}{3,600}\left(\frac{5,000}{75}\right)^2 = 9.72\text{ ft}$$

$$S = \pi D(L + D/2) = 283 \text{ ft}^2$$

Because S falls between 97 and 2,110 ft^2, equation 4.20 can be used to calculate the cost per vessel, C_v (assuming 304 stainless steel construction). Thus:

$$C_v = 271\, S^{0.778} = \$21,900$$

Adsorber Equipment Cost: **Recall that the adsorber equipment cost is comprised of the adsorber vessels, carbon, and the condenser, decanter, fan, pumps and other equipment usually included in the adsorber price. The cost of the latter items are "factored" from the combined cost of the vessels and carbon. Combining equations 4.21 and 4.22, we have:**

$$C_A = 5.82Q^{-0.133}[C_c + (N_A + N_D)C_v]$$

Substitution of the above values yields:

$$C_A = \$149,300$$

Cost of Auxiliary Equipment: **Assume that costs for the following auxiliary equipment have been estimated from data in other parts of the *Manual*:**

Ductwork	**$16,500**
Dampers	**7,200**
Stack	**8,500**
Total	**$32,200**

Total Capital Investment: **The total capital investment is factored from the sum of the adsorber unit and auxiliary equipment cost, as displayed in Table 4.4. Note that no line item cost has been shown for instrumentation, for this cost is typically included in the adsorber price. Therefore:**

Purchased Equipment Cost = "B" = 1.08× "A"
= 1.08× ($149,300 + $32,200) = $196,000

Table 4.4: Capital Costs for Carbon Adsorber System Example Problem

Cost Item	Cost
Direct Costs	
Purchased equipment costs	
Adsorber vessels and carbon	$149,300
Auxiliary equipment	32,200
Sum = A	$181,500
Instrumentation, 0.1A[a]	—
Sales taxes, 0.03A	5,450
Freight, 0.05A	9,080
Purchased equipment cost, B	$196,000
Direct installation costs	
Foundation and supports, 0.08B	15,680
Handling & erection, 0.14B	27,440
Electrical, 0.04B	7,840
Piping, 0.02B	3,920
Insulation for ductwork, 0.01B	1,960
Painting, 0.01B	1,960
Direct installation cost	$58,800
Site preparation	—
Facilities and buildings	—
Total Direct Cost	$254,800
Indirect Costs (installation)	
Engineering, 0.10B	19,600
Construction and field expenses, 0.05B	9,800
Contractor fees, 0.10B	19,600
Start-up, 0.02B	3,920
Performance test, 0.01B	1,960
Contingencies, 0.03B	5,880
Total Indirect Cost	$60,760
Total Capital Investment (rounded)	$316,000

[a] The cost for this is included in the adsorber equipment cost.

And:

Total Capital Investment (rounded) = 1.61 × "B" = $316,000

Annual Costs: Table 4.5 gives the direct and indirect annual costs for the carbon adsorber system, as calculated from the factors in Section 4.4. Except for electricity, the calculations in the table show how these costs were derived. The following discussion will deal with the electricity cost.

First, recall that the electricity includes the power for the system fan, bed drying/cooling fan, and the cooling water pump. (The solvent pump motor is normally so small that its power consumption may be neglected.) These consumptions are calculated as follows:

- *System fan:* From equation 4.27:

$$\text{kWh}_{sf} = 0.746\text{kW/hp} \times 2.50 \times 10^{-4} Q \Delta P_s \times \theta_s$$

But:

$$\Delta P_s \text{ (inches water)} = \Delta P_b + 1 = t_b(0.03679 v_b + 1.107 \times 10^{-4} v_b^2) + 1$$

(The latter expression was derived from equation 4.25, assuming that the carbon used in this example system is Calgon's "PCB", 4 × 10 mesh size.)

By assuming a carbon bed density of 30 lb/ft^3, Equation 4.26 can be used to calculate the bed thickness (t_b):

$$\text{Bed thickness} = t_b = \frac{0.0333\, M_c'}{A_b} = \frac{0.0333\, M_c'}{LD} = 1.80 \text{ ft.}$$

Thus:

$$\Delta P_s = 1 + 1.80(0.03679 \times 75 + 1.107 \times 10^{-4} \times 75^2) = 7.09 \text{ inches}$$

and finally:

$$\text{kWh}_{sf} = 0.746 \times 2.50 \times 10^{-4} \times 7.09 \text{ in.} \times 10,000 \text{ acfm} \times 8640 \text{ h/yr}$$

$$\text{kWh}_{sf} = 114,200 \text{ kWh/yr}$$

Table 4.5: Annual Costs for Carbon Adsorber System Example Problem

Cost Item	Calculations	Cost
Direct Annual Costs, DC		
Operating Labor		
Operator	$\frac{0.5\text{ h}}{\text{shift}} \times \frac{3\text{ sh}}{\text{day}} \times \frac{360\text{ days}}{\text{yr}} \times \frac{\$12}{\text{h}}$	$6,480
Supervisor	15% of operator = .15 × 6,480	970
Operating materials		—
Maintenance		
Labor	$\frac{0.5\text{ h}}{\text{shift}} \times \frac{3\text{ sh}}{\text{day}} \times \frac{360\text{ days}}{\text{yr}} \times \frac{\$13.20}{\text{h}}$	7,130
Material	100% of maintenance labor	7,130
Replacement parts, carbon (5 year life)		
Replacement labor	0.2638 ($0.05/lb × 10,800 lb)	140
Carbon cost[a]	0.2638($21, 600 × 1.08)	6,150
Utilities		
Electricity	$0.06 /kWh × 131,000 kWh/yr	7,860
Steam	$\frac{3.5\text{ lb}}{\text{lb VOC}} \times \frac{\$6}{10^3\text{ lb}} \times \frac{100\text{ lb VOC}}{\text{h}} \times \frac{8640\text{ h}}{\text{yr}}$	18,140
Cooling water	$\frac{3.43\text{ gal}}{\text{lb steam}} \times \frac{(3.5 \times 100 \times 8640)\text{ lb steam}}{\text{yr}} \times \frac{\$0.20}{10^3\text{ gal}}$	2,070
Total DC		$56,070
Indirect Annual Costs, IC		
Overhead	60% of sum of operating, supv., & maint. labor & maint. materials = 0.6(6,480 + 970 + 7,130 + 7,130)	13,030
Administrative charges	2% of Total Capital Investment = 0.02($316,000)	6,320
Property tax	1% of Total Capital Investment = 0.01($316,000)	3,160
Insurance	1% of Total Capital Investment = 0.01($316,000)	3,160
Capital recovery[b]	0.1628[316,000 − 0.05(10,800) − 1.08(21,600)]	47,560
Total IC		$73,230
Recovery Credit (toluene)	.98($\frac{\$0.0553}{\text{lb}} \times \frac{8,640\text{ h}}{\text{yr}} \times \frac{100\text{ lb}}{\text{h}}$)	(46,820)
Total Annual Cost (rounded)		$82,500

[a] The 1.08 factor is for freight and sales taxes.

[b] The capital recovery cost factor, CRF, is a function of the adsorber or equipment life and the opportunity cost of the capital (*i.e.*, interest rate). For example, for a 10 year equipment life and a 10% interest rate, CRF = 0.1628.

- *Bed drying/cooling fan:* During the drying/cooling cycle, the pressure drop through the bed also equals ΔP_b. However, as section 4.4.1.3 indicates, the flow and operating time are different. For the air flow, take the midpoint of the range given on page 4-30 (100 ft^3 air/lb carbon) and divide by 2 hours (the bed drying/cooling time), yielding: 100 ft^3/lb × 3,600 lbs × 1/120 min = 3,000 acfm. Substituting this into equation 4.27 results in:

  $$2.50 \times 10^{-4} \times 7.09 \text{ inches} \times 3,000 \text{ acfm} = 5.32 \text{ hp}$$

  From equation 4.28, we get:

  $$\theta_{cf} = (0.4)(5 \text{ h})(2)(8,640 \text{ h})/12 \text{ h} = 2,880 \text{ h}$$

  Thus:

  $$\text{kWh}_{cf} = 0.746 \text{ kW/hp} \times 5.32 \text{ hp} \times 2,880 \text{ h} = 11,400 \text{ kWh/yr}$$

- *Cooling water pump:* The cooling water pump horsepower is calculated from equation 4.29. Here, let $\eta = 63\%$ and H = 100 ft. The cooling water flow (q_{cw}) is the quotient of the annual cooling water requirement and the annual pump operating time. From the data in Table 4.5, we obtain the cooling water requirement: 10,400,000 gal/yr. The pump annual operating time is obtained from equation 4.28 (substituting 0.6 for 0.4), or $\theta_{cwp} = (0.6)(5 \text{ h})(2)(8,640)/12 =$ 4,320 h/yr.

  Thus:

  $$\text{hp}_{cwp} = \frac{(2.52 \times 10^{-4})(100\text{ft})}{0.63} \times \frac{10,400,000 \text{ gal/yr}}{4,320 \text{ h/yr} \times 60 \text{ min/yr}} = 1.60 \text{ hp}$$

  And:

  $$\text{kWh}_{cwp} = 0.746 \text{ kW/h} \times 1.60 \text{ hp} \times 4,320 \text{ h/yr} = 5,160 \text{ kWh/yr}$$

  Summing the individual power consumptions, we get the value shown in Table 4.5: 131,000 kWh/yr.

Recovery Credit: As Table 4.5 indicates, a credit for the recovered toluene has been taken. However, to account for miscellaneous losses and contamination, the toluene is arbitrarily valued at one-half the current (November 1989) market price (\$0.0533/lb = \$111/ton).[15]

Total Annual Cost: The sum of the direct and indirect annual costs, less the toluene recovery credit, yields a net total annual cost of $82,500. Clearly, this "bottom line" is very sensitive to the recovery credit and, in turn, the value given the recovered toluene. For instance, if it had been valued at the full market price ($221/ton), the credit would have doubled and the total annual cost would have been a net credit of $35,700. Thus when incorporating recovery credits, it is imperative to select the value of the recovered product carefully.

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Chapter 5

FABRIC FILTERS

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Contents

5.1 Process Description 5-4
  5.1.1 Introduction 5-4
  5.1.2 Types of Fabric Filters 5-5
    5.1.2.1 Shaker Cleaning 5-5
    5.1.2.2 Reverse-air Cleaning 5-6
    5.1.2.3 Pulse-jet Cleaning 5-7
  5.1.3 Auxiliary Equipment 5-8
  5.1.4 Fabric Filtration Theory 5-9
    5.1.4.1 Reverse Air/Shake Deflate Baghouses 5-10
    5.1.4.2 Pulse-Jet Baghouses 5-13
5.2 Design Procedures 5-16
  5.2.1 Gas-to-Cloth Ratio 5-16
    5.2.1.1 Gas-to-Cloth Ratio From Similar Applications 5-16
    5.2.1.2 Gas-to-Cloth Ratio From Manufacturer's Methods 5-19
    5.2.1.3 Gas-to-Cloth Ratio From Theoretical/Empirical Equations 5-19
  5.2.2 Pressure Drop 5-23
  5.2.3 Particle Characteristics 5-24
  5.2.4 Gas Stream Characteristics 5-24
    5.2.4.1 Temperature 5-25

5.2.4.2 Pressure 5-25

5.2.5 Equipment Design Considerations 5-27

5.2.5.1 Pressure or Suction Housings 5-27

5.2.5.2 Standard or Custom Construction 5-27

5.2.5.3 Filter Media 5-28

5.3 Estimating Total Capital Investment 5-29

5.3.1 Equipment Cost 5-29

5.3.1.1 Bare Baghouse Costs 5-29

5.3.1.2 Bag Costs 5-35

5.3.1.3 Auxiliary Equipment 5-38

5.3.2 Total Purchased Cost 5-38

5.3.3 Total Capital Investment 5-38

5.4 Estimating Total Annual Costs 5-39

5.4.1 Direct Annual Cost 5-39

5.4.1.1 Operating and Supervisory Labor 5-41

5.4.1.2 Operating Materials 5-41

5.4.1.3 Maintenance 5-41

5.4.1.4 Replacement Parts 5-41

5.4.1.5 Electricity 5-42

5.4.1.6 Fuel 5-43

5.4.1.7 Water 5-43

5.4.1.8 Compressed Air 5-44

5.4.1.9 Dust Disposal 5-44

5.4.2 Indirect Annual Cost 5-44

5.4.3 Recovery Credits 5-45

5.4.4 Total Annual Cost 5-45

5.4.5 Example Problem 5-45

5.5 Acknowledgment 5-51

References 5-52

5.1 Process Description

5.1.1 Introduction

A fabric filter unit consists of one or more isolated compartments containing rows of fabric filter bags or tubes. Particle-laden gas passes up (usually) along the surface of the bags then radially through the fabric. Particles are retained on the upstream face of the bags, while the cleaned gas stream is vented to the atmosphere. The filter is operated cyclically alternating between relatively long periods of filtering and short periods of cleaning. During cleaning, dust that has accumulated on the bags is removed from the fabric surface and deposited in a hopper for subsequent disposal.

Fabric filters will collect particle sizes ranging from submicron to several hundred microns in diameter at efficiencies generally in excess of 99 or 99.9 percent. The dust cake collected on the fabric is primarily responsible for such high efficiency. Gas temperatures up to about 500°F, with surges to about 550°F can be accommodated routinely. Most of the energy used to operate the system appears as pressure drop across the bags and associated hardware and ducting. Typical values of pressure drop range from about 5 to 20 inches of water. Fabric filters are used where high-efficiency particle collection is required. Limitations are imposed by gas characteristics (temperature and corrosivity) and particle characteristics (primarily stickiness) that affect the fabric or its operation and that cannot be economically

accommodated.

Important process variables include particle characteristics, gas characteristics, and fabric properties. The most important design parameter is the air- or gas-to-cloth ratio, and the usual operating parameter of interest is pressure drop across the filter system. The major operating feature of fabric filters that distinguishes them from other gas filters is the ability to renew the filtering surface periodically by cleaning.

Another type of fabric filter currently being developed is the electrostatically enhanced filter. Pilot plant baghouses employing this new technology have shown substantially lower pressure drops than conventional filter designs. Further, some cost analyses have shown that electrostatically enhanced baghouses could have lower lifetime costs than convention baghouses. The purpose of this chapter, however, is to focus only on currently available commercial filters. Anyone interested in electrostatically enhanced filtration may consult such references as Van Osdell et al [1], Viner et al [2], or Donovan [3].

In this section, the types of fabric filters and the auxiliary equipment required are discussed first from a general viewpoint. Then, fabric filtration theory as applied to each type of filter is discussed to lay a foundation for the sizing procedures outlined in the Section 5.2.

5.1.2 Types of Fabric Filters

Fabric filters can be categorized by several means, including type of cleaning (shaker, reverse-air, pulse-jet), direction of gas flow (from inside the bag towards the outside or vice versa), location of the system fan (suction or pressure), or size (low, medium, or high gas flow quantity). Of these four approaches, the cleaning method is probably the most distinguishing feature. Fabric filters are discussed in this section based on the type of cleaning employed.

5.1.2.1 Shaker Cleaning

For any type of cleaning, enough energy must be imparted to the fabric to overcome the adhesion forces holding dust to the bag. In shaker cleaning, used with inside to outside gas flow, this is accomplished by suspending the bag from a motor-driven hook or framework that oscillates. Motion may be imparted to the bag in several ways, but the general effect is to create a sine wave along the fabric. As the fabric moves outward, accumulated dust on the surface moves with the fabric. When the fabric reaches the limit of its extension, the patches of dust have enough inertia to tear away from the fabric and descend to the hopper.

For small, single-compartment baghouses, a lever attached to the shaker mechanism may be operated manually at appropriate intervals, typically at the end of a shift. In multi-compartment baghouses, a timer or a pressure sensor responding to system pressure drop initiates bag shaking automatically. The compartments operate in sequence so that one compartment at a time is cleaned. Forward gas flow to the compartment is stopped, dust is allowed to settle, residual gas flow stops, and the shaker mechanism is switched on for several seconds to a minute or more. The settling and shaking periods may be repeated, then the compartment is brought back on-line for filtering. Many large-scale shaker systems employ a small amount of reverse air during the shaker cycle to assist cleaning by deflating the bags.

Parameters that affect cleaning include the amplitude and frequency of the shaking motion and the tension of the mounted bag. The first two parameters are part of the baghouse design and generally are not changed easily. The tension is set when bags are installed. Typical values are about 4 Hz for frequency and 2 to 3 inches for amplitude (half-stroke).[4] Some installations allow easy adjustment of bag tension, while others require that the bag be loosened and reclamped to its attaching thimble.

The vigorous action of shaker systems tends to stress the bags and requires heavier and more durable fabrics. In the United States, woven fabrics are used almost exclusively for shaker cleaning.[5] European practice allows the use of felted fabrics at somewhat higher filtering velocities.

5.1.2.2 Reverse-air Cleaning

When glass fiber fabrics were introduced, a gentler means of cleaning the bags was needed to prevent premature degradation. Reverse-air cleaning was developed as a less intensive way to impart energy to the bags. In this method, gas flow to the bags is stopped in the compartment being cleaned, and a reverse flow of air is directed through the bags. This reversal of gas flow gently collapses the bags, and dust is removed from the fabric surface by shear forces developed between the dust and fabric as the latter changes its contours. Other differences between reverse-air and shaker cleaning are the installation of metal caps as an integral part of the bag and sewn-in rings to prevent complete collapse of the bag, which may be 30 feet long, during cleaning. Without these rings, collected dust tends to choke the bag as the fabric collapses in on itself during inside-to-outside filtering. As with multi-compartment shaker baghouses, the same cycle takes place in reverse-air baghouses of stopping forward gas flow and allowing dust to settle before cleaning action begins.

The source of reverse air is generally a separate system fan capable of supplying clean, dry air for one or two compartments at a gas-to-cloth ratio similar to that of the forward gas flow.

5.1.2.3 Pulse-jet Cleaning

This form of cleaning uses compressed air to force a burst of air down through the bag and expand it violently. As with shaker baghouses, the fabric reaches its extension limit and the dust separates from the bag. In pulse jets, however, filtering gas flows are opposite in direction when compared with shaker or reverse-air baghouses. Bags are mounted on wire cages to prevent collapse while the dusty gas flows from outside the bag to the inside. Instead of attaching both ends of the bag to the baghouse structure, the bag and cage assembly generally is attached only at the top. The bottom end of the assembly tends to move in the turbulent gas flow and may contact other bags, which accelerates wear.

Although some pulse-jet baghouses are compartmented, most are not. Bags are cleaned one row at a time when a timer initiates the burst of cleaning air through a quick-opening valve. A pipe across each row of bags carries the compressed air. The pipe is pierced above each bag so that cleaning air exits directly downward into the bag. Some systems direct the air through a short venturi that is intended to entrain additional cleaning air. The pulse opposes and interrupts forward gas flow only for a few tenths of a second. However, the quick resumption of forward flow redeposits most of the dust back on the clean bag or on adjacent bags. An advantage of

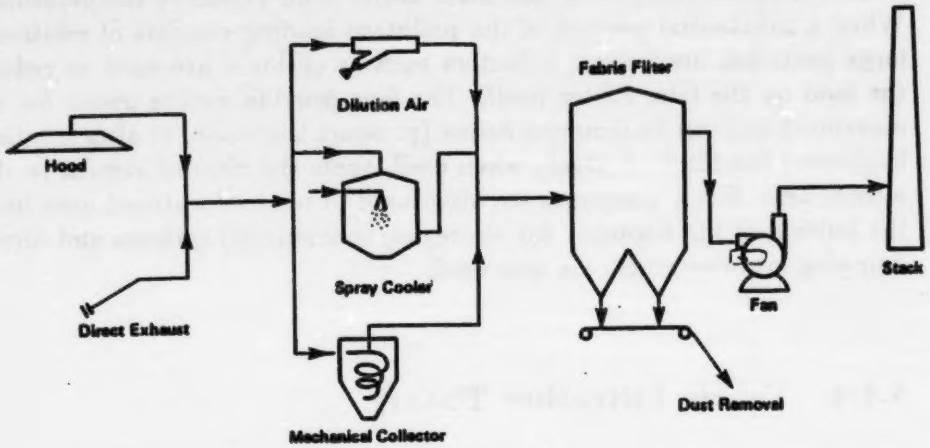


Figure 5.1: Typical alternative auxiliary equipment items used with fabric filter control systems.

pulse-jet cleaning is the reduction in baghouse size allowed by not having to build an extra compartment for off-line cleaning.

5.1.3 Auxiliary Equipment

The typical auxiliary equipment associated with fabric filter systems is shown in Figure 5.1. Along with the fabric filter itself, a control system typically includes the following auxiliary equipment: a capture device (*i.e.*, hood or direct exhaust connection); ductwork; dust removal equipment (screw conveyor, etc.); fans, motors, and starters; and a stack. In addition, spray chambers, mechanical collectors, and dilution air ports may be needed to precondition the gas before it reaches the fabric filter. Capture devices are usually hoods that exhaust pollutants into the ductwork or direct exhaust couplings attached to a process vessel. Hoods are more common, yet poorly designed hoods will allow pollutants to escape. Direct exhaust couplings are less common, requiring sweep air to be drawn through the process vessel, and may not be feasible in some processes. Ductwork provides a means of moving the exhaust stream to the control device. Spray chambers and dilution air ports are used to decrease the temperature of the pollutant stream to protect the filter fabric from excessive temperatures. When a substantial portion of the pollutant loading consists of relatively large particles, mechanical collectors such as cyclones are used to reduce the load on the fabric filter itself. The fans provide motive power for air movement and can be mounted before (pressure baghouse) or after (suction baghouse) the filter. A stack, when used, vents the cleaned stream to the atmosphere. Screw conveyors are often used to remove captured dust from the bottom of the hoppers. Air conveying (pneumatic) systems and direct dumping into containers are also used.

5.1.4 Fabric Filtration Theory

The key to designing a baghouse is to determine the face velocity that produces the optimum balance between pressure drop (operating cost) and baghouse size (capital cost). Major factors that affect design face velocity (or gas-to-cloth ratio), discussed in Section 5.2, include particle and fabric characteristics and gas temperature.

Although collection efficiency is another important measure of baghouse performance, it is generally assumed that a properly designed and well run baghouse will be highly efficient. Therefore, the design process focuses on the pressure drop.

There are several contributions to the pressure drop across a baghouse compartment, including the pressure drop from the flow through the inlet and outlet ducts, from flow through the hopper regions, and from flow through the bags. The pressure drop through the baghouse compartment (excluding the pressure drop across the bags) depends largely on the baghouse design and ranges from 1 to 2 inches of H_2O[3] in conventional designs and up to about 3 inches of H_2O in designs having complicated gas flow paths. This loss can be kept to a minimum (*i.e.*, 1 inch of H_2O or less) by investing in a flow modeling study of the proposed design. A study of this sort would cost on the order of $50,000 (in 1986). The pressure drop across the bags (also called the tube-sheet pressure drop) can be as high as 10 inches of H_2O or more. The tube-sheet pressure drop is a complex function of the physical properties of the dust and the fabric and the manner in which the baghouse is designed and operated. The duct and hopper losses are constant and can be minimized effectively through proper design based on a knowledge of the flow through the baghouse.[1]

Fabric filtration is inherently a batch process that has been adapted to continuous operation through clever engineering. One requirement for a continuously operating baghouse is that the dust collected on the bags must be removed periodically. Shaker and reverse-air baghouses are similar in the sense that they both normally use woven fabric bags, run at relatively low face velocities, and the filtration mechanism is cake filtration. That is, the fabric merely serves as a substrate for the formation of a dust cake that is the actual filtration medium. Pulse-jet baghouses generally use felt fabrics and run with a high gas-to-cloth ratio (about double that of shaker or reverse-air baghouses). The felt fabric may play a much more active role in the filtration process. This distinction between cake filtration and fabric filtration has important implications for the rate of pressure loss across the filter bags. The theoretical description of cake filtration is quite different from that for fabric filtration, and the design processes are quite different. Accurate fabric selection is aided by bench-scale filtration tests at less than one-tenth the cost of flow modeling. These tests can be used to investigate fabric effects on pressure drop, cake release during cleaning, and collection efficiency. Electrical properties of the fabric, such as resistivity and triboelectric order, may be measured to aid in fabric selection. Although their effects are generally poorly understood, electrical/electrostatic effects influence cake porosity and particle adhesion.[6,7,8]

The general equations used to size a baghouse follow beginning with the reverse air/shake deflate type of baghouse.

5.1.4.1 Reverse Air/Shake Deflate Baghouses

The construction of a baghouse begins with a set of specifications including average pressure drop, total gas flow, and other requirements; a maximum pressure drop may also be specified. Given these specifications, the designer must determine the maximum face velocity that can meet these requirements. The standard way to relate baghouse pressure drop to face velocity is given by the relation:

$$\Delta P(\theta) = S_{sys}(\theta)V_{ave} \qquad (5.1)$$

---

[1] A procedure for estimating duct pressure losses is given in the "Ductwork" chapter of this *Manual*.

where

$\Delta P(\theta)$ = the pressure drop across the filter, a function of time, θ (in. H_2O)

$S_{sys}(\theta)$ = system drag, a function of time [in. H_2O/(ft/min)]

V_{ave} = average (*i.e.*, design) face velocity, constant (ft/min)

For a multi-compartment baghouse, the system drag, which accounts for most of the drag from the inlet flange to the outlet flange of the baghouse, is determined as a combination of resistances representative of several compartments. For the typical case where the pressure drop through each compartment is the same, it can be shown that:[16]

$$S_{sys}(\theta) = \left[\frac{1}{M}\sum_{i=1}^{M}\frac{1}{S_i(\theta)}\right]^{-1} = \frac{1}{\frac{1}{M}\sum_{i=1}^{M}\frac{1}{S_i(\theta)}} = \frac{M}{\sum_{i=1}^{M}\frac{1}{S_i(\theta)}} \qquad (5.2)$$

where

M = number of compartments in the baghouse

$S_i(\theta)$ = drag across compartment *i*

The compartment drag is a function of the amount of dust collected on the bags in that compartment. In general, the dust will be distributed in a very nonuniform manner. That is, there will be a variation of dust load from one bag to the next and within a given bag there will also be a variation of dust load from one area to another. For a sufficiently small area *j* within compartment *i*, it can be assumed that the drag is a linear function of dust load:

$$S_{i,j}(\theta) = S_e + K_2 W_{i,j}(\theta) \qquad (5.3)$$

where

S_e = drag of a dust-free filter bag [in. H_2O/(ft/min)]

K_2 = dust cake flow resistance {[in. H_2O/(ft/min)]/(lb/ft^2)}

$W_{i,j}(\theta)$ = dust mass per unit area of area *j* in compartment *i*, "areal density" (lb/ft^2)

If there are N different areas of equal size within compartment *i*, each with a different drag $S_{i,j}$, then the total drag for compartment *i* can be computed in a manner analogous to Equation 5.2:

$$S_i(\theta) = \frac{N}{\sum \frac{1}{S_{i,j}(\theta)}}. \tag{5.4}$$

The constants S_e and K_2 depend upon the fabric and the nature and size of the dust. The relationships between these constants and the dust and fabric properties are not understood well enough to permit accurate predictions and so must be determined empirically, either from prior experience with the dust/fabric combination or from laboratory measurements. The dust mass as a function of time is defined as:

$$W_{i,j}(\theta) = W_r + \int_0^\theta C_{in}\, V_{i,j}(\theta)\, d\theta \tag{5.5}$$

where

W_r = dust mass per unit area remaining on a "clean" bag (lb/ft^2)
C_{in} = dust concentration in the inlet gas (lb/ft^3)
$V_{i,j}(\theta)$ = face velocity through area j of compartment i (ft/min)

It is assumed that the inlet dust concentration and the filter area are constant. The face velocity (gas-to-cloth ratio) through each filter area j and compartment i changes with time, starting at a maximum value just after cleaning and steadily decreasing as dust builds up on the bags. The individual compartment face velocities are related to the average face velocity by the expression:

$$\begin{aligned} V_{ave} &= \frac{\sum_i \sum_j V_{i,j}(\theta)\, A_{i,j}}{\sum_i \sum_j A_{i,j}} \qquad (5.6) \\ &= \frac{\sum_i \sum_j V_{i,j}}{M} \text{ (for M compartments with equal area)} \end{aligned}$$

Equations 5.1 through 5.6 reveal that there is no explicit relationship between the design face velocity and the tube-sheet pressure drop. On the contrary, the pressure drop that results from a given design can only be determined by the simultaneous solution of Equations 5.1 through 5.5, with Equation 5.6 as a constraint on that solution. This conclusion has several implications for the design process. The design requires an iterative procedure: one must begin with a known target for the average pressure drop, propose a baghouse design (number of compartments, length of filtration period, etc.), assume a face velocity that will yield that pressure drop, and solve the system of Equations 5.1 through 5.6 to verify that the calculated pressure drop equals the target pressure drop. This procedure is repeated until the specified face velocity yields an average pressure drop (and maximum pressure drop, if applicable) that is sufficiently close to the design specification.

5.1.4.2 Pulse-Jet Baghouses

The distinction between pulse-jet baghouses and reverse-air and shaker baghouses is basically the difference between cake filtration and composite dust/fabric filtration (noncake filtration). This distinction is more a matter of convenience than physics. In reality, pulse-jet baghouses have been designed to operate in a variety of modes. Some pulse jets remain on-line at all times and are cleaned frequently. Others are taken off-line for cleaning at relatively long intervals. Obviously, if a compartment remains on-line long enough without being cleaned, then the filtration mechanism becomes that of cake filtration. A complete model of pulse-jet filtration therefore must account for the depth filtration occurring on a relatively clean pulse-jet filter, the cake filtration that inevitably results from prolonged periods on-line, and the transition period between the two regimes.

Besides the question of filtration mechanism, there is also the question of cleaning method. If a compartment is taken off-line for cleaning, then the dust that is removed from the bags will fall into the dust hopper before forward gas flow resumes. If a compartment is cleaned while on-line, then only a small fraction of the dust removed from the bag will fall to the hopper. The remainder of the dislodged dust will be redeposited (*i.e.*, "recycled") on the bag by the forward gas flow. The redeposited dust layer has different pressure drop characteristics than the freshly deposited dust. The modeling work that has been done to date focuses on the on-line cleaning method. Dennis and Klemm[9] proposed the following model of drag across a pulse-jet filter:

$$S = S_e + (K_2)_c W_c + K_2 W_o \qquad (5.7)$$

where

S = drag across the filter
S_e = drag of a just-cleaned filter
$(K_2)_c$ = specific dust resistance of the recycling dust
W_c = areal density of the recycling dust
K_2 = specific dust resistance of the freshly deposited dust
W_o = areal density of the freshly deposited dust

This model has the advantage that it can easily account for all three regimes of filtration in a pulse-jet baghouse. As in Equations 5.1 to 5.6, the drag, filtration velocity and areal densities are functions of time, θ. For given operating conditions, however, the values of S_e, $(K_2)_c$, and W_c may be assumed to be constant, so that they can be grouped together:

$$\Delta P = (PE)_{\Delta w} + K_2 W_o V \tag{5.8}$$

where

ΔP = pressure drop (in. H_2O)
V = filtration velocity (ft/min)
$(PE)_{\Delta w}$ = $[S_e + (K_2)_c W_c]\ V$

Equation 5.8 describes the pressure drop behavior of an individual bag. To extend this single bag result to a multiple-bag compartment, Equation 5.7 would be used to determine the individual bag drag and the total baghouse drag would then be computed as the sum of the parallel resistances. Pressure drop would then be calculated as in Equation 5.1. It seems reasonable to extend this analysis to the case where the dust is distributed unevenly on the bag and then apply Equation 5.7 to each area on the bag, followed by an equation analogous to 5.4 to compute the overall bag drag. The difficulty in doing this is that one must assume values for W_c for each different area to be modeled.

The disadvantage of the model represented by Equations 5.7 and 5.8 is that the constants, S_e, K_2, and W_c, cannot be predicted at this time. Consequently, correlations of laboratory data must be used to determine the value of $(PE)_{\Delta w}$. For the fabric-dust combination of Dacron felt and coal fly ash, Dennis and Klemm[9] developed an empirical relationship between $(PE)_{\Delta w}$, the face velocity, and the cleaning pulse pressure. This relationship (converted from metric to English units) is as follows:

$$(PE)_{\Delta w} = 6.08 V_f P_j^{-0.65} \qquad (5.9)$$

where

V_f = face velocity (ft/min)
P_j = pressure of the cleaning pulse (usually 60 to 100 psig; see Section 5.4.1)

This equation is essentially a regression fit to a limited amount of laboratory data and should not be applied to other dust/fabric combinations. The power law form of Equation 5.9 may not be valid for other dusts or fabrics. Consequently, more data should be collected and analyzed before the model represented by Equation 5.9 can be used for rigorous sizing purposes.

Another model that shows promise in the prediction of noncake filtration pressure drop is that of Leith and Ellenbecker[10] as modified by Koehler and Leith.[11] In this model, the tube-sheet pressure drop is a function of the clean fabric drag, the system hardware, and the cleaning energy. Specifically:

$$\Delta P = \frac{1}{2}\left[P_s + K_1 V_f - \sqrt{(P_s - K_1 V_f)^2 - 4W_o K_2/K_3}\right] + K_v V_f^2 \qquad (5.10)$$

where

P_s = maximum static pressure achieved in the bag during cleaning
K_1 = clean fabric resistance
V_f = face velocity
K_2 = dust deposit flow resistance
K_3 = bag cleaning efficiency coefficient
K_v = loss coefficient for the venturi at the inlet to the bag

Comparisons of laboratory data with pressure drops computed from Equation 5.10 [10,11] are in close agreement for a variety of dust/fabric combinations. The disadvantage of Equation 5.10 is that the constants K_1, K_2, and K_3 must be determined from laboratory measurements. The most difficult one to determine is the K_3 value, which can only be found by making measurements in a pilot-scale pulse-jet baghouse. A limitation of laboratory measurements is that actual filtration conditions cannot always be adequately simulated. For example, a redispersed dust may not have the same size distribution as the original dust, thereby yielding different values of K_1, K_2, and K_3 than would be measured in an operating baghouse.

5.2 Design Procedures

5.2.1 Gas-to-Cloth Ratio

The gas-to-cloth ratio is difficult to estimate from first principles. However, shortcut methods of varying complexity allow rapid estimation. Descriptions of three methods of increasing difficulty follow. For shaker and reverse-air baghouses, the third method is best performed with publicly available computer model programs.

The methods outlined below pertain to conventional baghouses. Use of electrostatic stimulation (an emerging technology) allows a higher gas-to-cloth ratio at a given pressure drop; thus a smaller baghouse structure and fewer bags are needed. This reduces total annual cost by up to 30%. Viner and Locke [13] discuss cost and performance models for electrostatically stimulated fabric filters.

5.2.1.1 Gas-to-Cloth Ratio From Similar Applications

After a fabric has been selected, the gas-to-cloth ratio can be determined using Table 5.1. Column 1 shows the type of dust; column 2 shows the gas-to-cloth ratios for woven fabric; and column 3 shows gas-to-cloth ratios for felted fabrics. Notice that these values are all *net* gas-to-cloth ratios. The net gas-to-cloth ratio is equal to the total actual volumetric flow rate in cubic feet per minute divided by the net cloth area in square feet. This ratio, in units of feet per minute, affects pressure drop and bag life. The net cloth area is determined by dividing the gas-to-cloth ratio into the actual cubic feet per minute flow of the exhaust gas stream. For an intermittent-type baghouse that is shut down for cleaning, this is the total, or gross, cloth area. However, for continuously operated filters, the area must be increased to allow the shutting down of one or more compartments for cleaning. Table 5.2 provides a guide for adjusting the net area to the gross area, which determines the size of a continuously cleaned filter.

Table 5.1: Gas-to-Cloth Ratios[*,a]
(actual $ft^3/min)/(ft^2$ of net cloth area)

Dust	Shaker/Woven Reverse-Air/Woven	Pulse Jet/Felt Reverse-Air Felt
Alumina	2.5	8
Asbestos	3.0	10
Bauxite	2.5	8
Carbon Black	1.5	5
Coal	2.5	8
Cocoa, Chocolate	2.8	12
Clay	2.5	9
Cement	2.0	8
Cosmetics	1.5	10
Enamel Frit	2.5	9
Feeds, Grain	3.5	14
Feldspar	2.2	9
Fertilizer	3.0	8
Flour	3.0	12
Fly Ash	2.5	5
Graphite	2.0	5
Gypsum	2.0	10
Iron Ore	3.0	11
Iron Oxide	2.5	7
Iron Sulfate	2.0	6
Lead Oxide	2.0	6
Leather Dust	3.5	12
Lime	2.5	10
Limestone	2.7	8
Mica	2.7	9
Paint Pigments	2.5	7
Paper	3.5	10
Plastics	2.5	7
Quartz	2.8	9
Rock Dust	3.0	9
Sand	2.5	10
Sawdust (Wood)	3.5	12
Silica	2.5	7
Slate	3.5	12
Soap, Detergents	2.0	5
Spices	2.7	10
Starch	3.0	8
Sugar	2.0	7
Talc	2.5	10
Tobacco	3.5	13
Zinc Oxide	2.0	5

[*] Reference [12]

[a] Generally safe design values; application requires consideration of particle size and grain loading.

Table 5.2: Approximate Guide to Estimate Gross Cloth Area *

Net Cloth Area (ft^2)	Gross Cloth Area (ft^2)	
1-4,000	Multiply by	2
4,001-12,000	"	1.5
12,001-24,000	"	1.25
24,001-36,000	"	1.17
36,001-48,000	"	1.125
48,001-60,000	"	1.11
60,001-72,000	"	1.10
72,001-84,000	"	1.09
84,001-96,000	"	1.08
96,001-108,000	"	1.07
108,001-132,000	"	1.06
132,001-180,000	"	1.05
above 180,001	"	1.04

*Reference [14]

5.2.1.2 Gas-to-Cloth Ratio From Manufacturer's Methods

Manufacturers have developed nomographs and charts that allow rapid estimation of the gas-to-cloth ratio. Two examples are given below, one for shaker-cleaned baghouses and the other for pulse-jet cleaned baghouses.

For shaker baghouses, Table 5.3 gives a factor method for estimating the ratio. Ratios for several materials in different operations are presented, but are modified by factors for particle size and dust load. Directions and an example are included. Gas-to-cloth ratios for reverse-air baghouses would be about the same or a little more conservative compared to the Table 5.3 values.

For pulse-jet baghouses, which normally operate at two or more times the gas-to-cloth ratio of reverse-air baghouses, another factor method[15] has been modified with equations to represent temperature, particle size, and dust load:

$$V = 2.878\, A\, B\, T^{-0.2335}\, L^{-0.06021}\, (0.7471 + 0.0853 \ln D) \qquad (5.11)$$

where

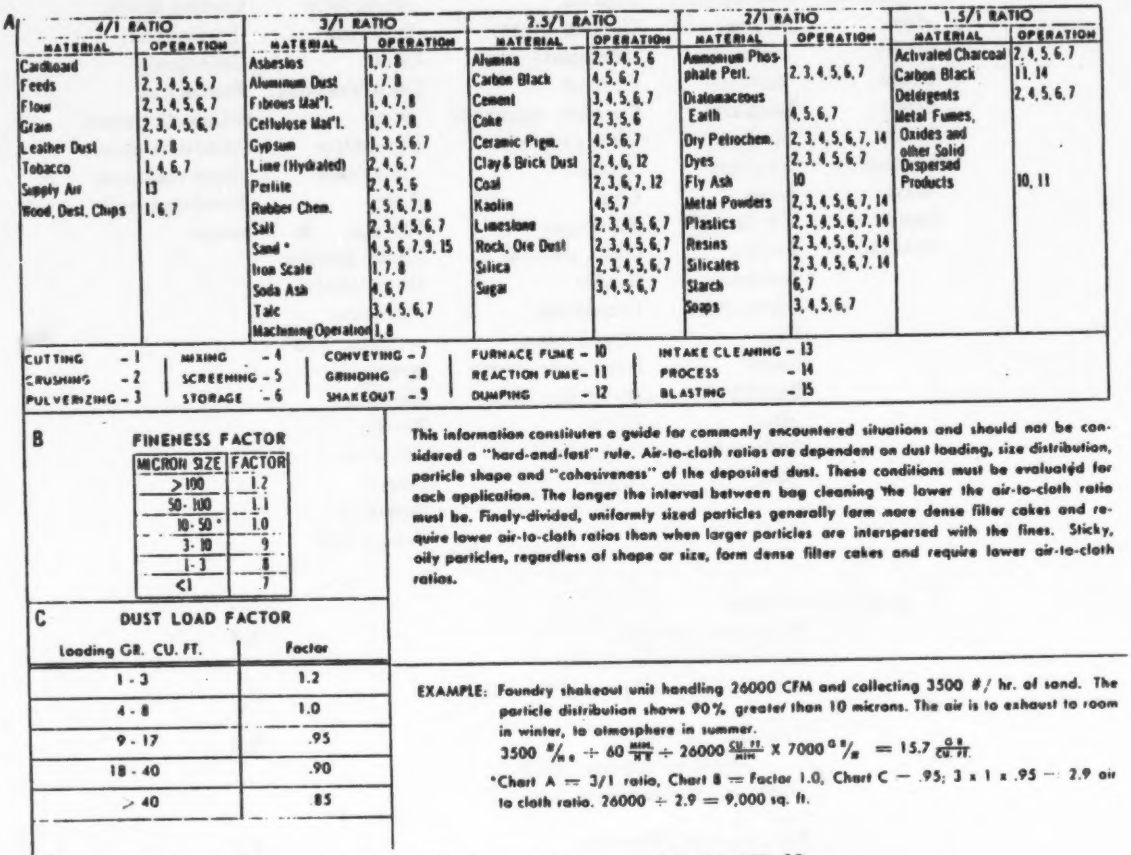
V = gas-to-cloth ratio (ft/min)
A = material factor, from Table 5.4
B = application factor, from Table 5.4
T = temperature, (°F, between 50 and 275)
L = inlet dust loading (gr/ft^3, between 0.05 and 100)
D = mass mean diameter of particle (μm, between 3 and 100)

For temperatures below 50°F, use T = 50 but expect decreased accuracy; for temperatures above 275°F, use T = 275. For particle mass mean diameters less than 3 μm, the value of D is 0.8, and for diameters greater than 100 μm, D is 1.2. For dust loading less than 0.05 gr/ft^3, use L = 0.05; for dust loading above 100 gr/ft^3, use L = 100.

5.2.1.3 Gas-to-Cloth Ratio From Theoretical/Empirical Equations

Shaker and reverse-air baghouses The system described by Equations 5.1 through 5.6 is complicated; however, numerical methods can be used to

Table 5.3: Manufacturer's Factor Method for Estimating Gas-to-Cloth Ratios for Shaker Baghouses



A

4/1 RATIO		3/1 RATIO		2.5/1 RATIO		2/1 RATIO		1.5/1 RATIO	
MATERIAL	OPERATION	MATERIAL	OPERATION	MATERIAL	OPERATION	MATERIAL	OPERATION	MATERIAL	OPERATION
Cardboard	1	Asbestos	1, 7, 8	Alumina	2, 3, 4, 5, 6	Ammonium Phosphate Fert.	2, 3, 4, 5, 6, 7	Activated Charcoal	2, 4, 5, 6, 7
Feeds	2, 3, 4, 5, 6, 7	Aluminum Dust	1, 7, 8	Carbon Black	4, 5, 6, 7	Diatomaceous Earth	4, 5, 6, 7	Carbon Black	11, 14
Flour	2, 3, 4, 5, 6, 7	Fibrous Mat'l.	1, 4, 7, 8	Cement	3, 4, 5, 6, 7	Dry Petrochem.	2, 3, 4, 5, 6, 7, 14	Detergents	2, 4, 5, 6, 7
Grain	2, 3, 4, 5, 6, 7	Cellulose Mat'l.	1, 4, 7, 8	Coke	2, 3, 5, 6	Dyes	2, 3, 4, 5, 6, 7	Metal Fumes, Oxides and other Solid Dispersed Products	10, 11
Leather Dust	1, 7, 8	Gypsum	1, 3, 5, 6, 7	Ceramic Pigm.	4, 5, 6, 7	Fly Ash	10		
Tobacco	1, 4, 6, 7	Lime (Hydrated)	2, 4, 6, 7	Clay & Brick Dust	2, 4, 6, 12	Metal Powders	2, 3, 4, 5, 6, 7, 14		
Supply Air	13	Perlite	2, 4, 5, 6	Coal	2, 3, 6, 7, 12	Plastics	2, 3, 4, 5, 6, 7, 14		
Wood, Dust, Chips	1, 6, 7	Rubber Chem.	4, 5, 6, 7, 8	Kaolin	4, 5, 7	Resins	2, 3, 4, 5, 6, 7, 14		
		Salt	2, 3, 4, 5, 6, 7	Limestone	2, 3, 4, 5, 6, 7	Silicates	2, 3, 4, 5, 6, 7, 14		
		Sand *	4, 5, 6, 7, 9, 15	Rock, Ore Dust	2, 3, 4, 5, 6, 7	Starch	6, 7		
		Iron Scale	1, 7, 8	Silica	2, 3, 4, 5, 6, 7	Soaps	3, 4, 5, 6, 7		
		Soda Ash	4, 6, 7	Sugar	3, 4, 5, 6, 7				
		Talc	3, 4, 5, 6, 7						
		Machining Operation	1, 8						

CUTTING - 1
CRUSHING - 2
PULVERIZING - 3
MIXING - 4
SCREENING - 5
STORAGE - 6
CONVEYING - 7
GRINDING - 8
SHAKEOUT - 9
FURNACE FUME - 10
REACTION FUME - 11
DUMPING - 12
INTAKE CLEANING - 13
PROCESS - 14
BLASTING - 15

B

FINENESS FACTOR

MICRON SIZE	FACTOR
> 100	1.2
50 - 100	1.1
10 - 50 *	1.0
3 - 10	.9
1 - 3	.8
<1	.7

C

DUST LOAD FACTOR

Loading GR. CU. FT.	Factor
1 - 3	1.2
4 - 8	1.0
9 - 17	.95
18 - 40	.90
> 40	.85

This information constitutes a guide for commonly encountered situations and should not be considered a "hard-and-fast" rule. Air-to-cloth ratios are dependent on dust loading, size distribution, particle shape and "cohesiveness" of the deposited dust. These conditions must be evaluated for each application. The longer the interval between bag cleaning the lower the air-to-cloth ratio must be. Finely-divided, uniformly sized particles generally form more dense filter cakes and require lower air-to-cloth ratios than when larger particles are interspersed with the fines. Sticky, oily particles, regardless of shape or size, form dense filter cakes and require lower air-to-cloth ratios.

EXAMPLE: Foundry shakeout unit handling 26000 CFM and collecting 3500 #/ hr. of sand. The particle distribution shows 90% greater than 10 microns. The air is to exhaust to room in winter, to atmosphere in summer.

$$3500 \frac{\#}{HR} \div 60 \frac{MIN}{HR} \div 26000 \frac{CU.\ FT.}{MIN} \times 7000 \frac{GR}{\#} = 15.7 \frac{GR}{CU.\ FT.}$$

*Chart A = 3/1 ratio, Chart B = Factor 1.0, Chart C = .95; 3 x 1 x .95 = 2.9 air to cloth ratio. 26000 ÷ 2.9 = 9,000 sq. ft.

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Table 5.4: Factors for Pulse-Jet Gas-to-Cloth Ratios [*]

A. Material Factor

15[a]	12	10	9.0	6.0[b]
Cake mix Cardboard dust Cocoa Feeds Flour Grain Leather dust Sawdust Tobacco	Asbestos Buffing dust Fibrous and cellulosic material Foundry shakeout Gypsum Lime (hydrated) Perlite Rubber chemicals Salt Sand Sandblast dust Soda ash Talc	Alumina Aspirin Carbon black (finished) Cement Ceramic pigments Clay and brick dusts Coal Fluorspar Gum, natural Kaolin Limestone Perchlorates Rock dust, ores and minerals Silica Sorbic acid Sugar	Ammonium phosphate-fertiliser Cake Diatomaceous earth Dry petrochemicals Dyes Fly ash Metal powder Metal oxides Pigments metallic and synthetic Plastics Resins Silicates Starch Stearates Tannic acid	Activated carbon Carbon black (molecular) Detergents Fumes and other dispersed products direct from reactions Powdered milk Soaps

B. Application Factor

Nuisance Venting Relief of transfer points, conveyors, packing stations, etc.	1.0
Product Collection Air conveying-venting, mills, flash driers, classifiers, etc.	0.9
Process Gas Filtration Spray driers, kilns, reactors, etc.	0.8

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[*] Reference [15]

[a] In general, physically and chemically stable materials.

[b] Also includes those solids that are unstable in their physical or chemical state due to hygroscopic nature, sublimation, and/or polymerization.

obtain an accurate solution. A critical weakness in baghouse modeling that has yet to be overcome is the lack of a fundamental description of the bag cleaning process. That is, to solve Equations 5.1 through 5.6, the value of W_r (the dust load after cleaning) must be known. Clearly, there must be a relationship between the amount and type of cleaning energy and the degree of dust removal from a bag. Dennis *et al.*[16] have developed correlations for the removal of coal fly ash from woven fiberglass bags by shaker cleaning and by reverse air cleaning. These correlations have been incorporated into a computer program that generates the solution to the above system of equations.[9,17,18] If one were to apply the correlations developed with coal ash and woven glass fabrics to other dust/fabric combinations, the accuracy of the results would depend on how closely that dust/fabric combination mimicked the coal ash/woven glass fabric system.

Physical factors that affect the correlation include the particle size distribution, adhesion and electrostatic properties of the dust and fabric, and fabric weave, as well as cleaning energy. More research is needed in this area of fabric filtration.

The rigorous design of a baghouse thus involves several steps. First, the design goal for average pressure drop (and maximum pressure drop, if necessary) must be specified along with total gas flow rate and other parameters, such as S_e and K_2 (obtained either from field or laboratory measurements). Second, a face velocity is assumed and the number of compartments in the baghouse is computed based on the total gas flow, face velocity, bag size, and number of bags per compartment. (Typical compartments in the U.S. electric utility industry use bags 1 ft in diameter by 30 ft long with 400 bags per compartment.) Standard practice is to design a baghouse to meet the specified pressure drop when one compartment is off-line for maintenance. The third step is to specify the operating characteristics of the baghouse (*i.e.*, filtration period, cleaning period, and cleaning mechanism). Fourth, the designer must specify the cleaning efficiency so that the residual dust load can be estimated. Finally, the specified baghouse design is used to establish the details for Equations 5.1 through 5.6, which are then solved numerically to establish the pressure drop as a function of time. The average pressure drop is then computed by integrating the instantaneous pressure drop over the filtration cycle and dividing by the cycle time. If the computed average is higher than the design specification, then the face velocity must be reduced and the procedure repeated. If the computed average pressure drop is significantly lower than the design specification, then the proposed baghouse was oversized and should be made smaller by increasing the face velocity and repeating the procedure. When the computed average pressure drop comes sufficiently close to the assumed specified value, then the design has been determined. A complete description of the modeling process can be found in the reports by Dennis *et al.*[16,18] A critique on the accuracy of the model is presented by Viner *et al.*[19]

Pulse-jet baghouses The overall process of designing a pulse jet baghouse is actually simpler than that required for a reverse-air or shaker baghouse if the baghouse remains on-line for cleaning. The first step is to specify what the desired average tube-sheet pressure drop should be. Second, the operating characteristics of the baghouse must be established (*e.g.*, on-line time, cleaning energy). Third, the designer must obtain values for the coefficients in either Equation 5.9 or Equation 5.10 from field, pilot plant, or laboratory measurements. Fourth, a value is estimated for the face velocity and the appropriate equation (Equation 5.8 or 5.10) is solved for the pressure drop as a function of time for the duration of the filtration cycle. This information is used to calculate the cycle average pressure drop. If the calculated pressure drop matches the specified pressure drop, then the procedure is finished. If not, then the designer must adjust the face velocity and repeat the procedure.

5.2.2 Pressure Drop

Pressure drop for the bags can be calculated rigorously from the equations given in the preceding section if values for the various parameters are known. Frequently they are not known. For quick estimation, a maximum pressure drop of 5 to 10 in. H_2O across the baghouse and 10 to 20 in. H_2O across the entire system can be assumed if it contains much ductwork.

A comparable form of Equations 5.1 and 5.3 that may be used for pressure drop across the fabric in a shaker or reverse-air baghouse is:

$$\Delta P = S_e V + K_2 C_i V^2 \theta \tag{5.12}$$

where

ΔP = pressure drop (in. H_2O)
S_e = effective residual drag of the fabric [in. $H_2O/(ft/min)$]
V = superficial face velocity or gas-to-cloth ratio (ft/min)
K_2 = specific resistance coefficient of the dust {[in. $H_2O/(ft/min)]/(lb/ft^2)$}
C_i = inlet dust concentration (lb/ft^3)
θ = filtration time (min)

Although there is much variability, values for S_e may range from about 0.2 to 2 in. $H_2O/(ft/min)$ and for K_2 from 1-2 to 30-40 [in. $H_2O/(ft/min)]/(lb/ft^2)$. Typical values for coal fly ash are about 1 to 4. Inlet concentrations vary from less than 0.05 gr/ft^3 to more than 100 gr/ft^3, but a more nearly typical range is from about 0.5 to 10 gr/ft^3. Filtration times may range from about 20 to 90 minutes for continuous duty baghouses, but 30 to 60 minutes is more frequently found. For pulse-jet baghouses, use Equations 5.8 and 5.9 to estimate ΔP, after substituting $C_iV\theta$ for W_o and $(PE)_{\Delta w}$ for S_eV.

5.2.3 Particle Characteristics

Particle size distribution and adhesiveness are the most important particle properties that affect design procedures. Smaller particle sizes can form a denser cake, which increases pressure drop. As shown in Table 5.3 and Equation 5.11, the effect of decreasing average particle size is a lower applicable gas-to-cloth ratio.

Adhering particles, such as oily residues or electrostatically active plastics, may require installing equipment that injects a precoating material onto the bag surface, which acts as a buffer that traps the particles and prevents them from blinding or permanently plugging the fabric pores. Informed fabric selection may eliminate electrostatic problems.

5.2.4 Gas Stream Characteristics

Moisture and corrosives content are the major gas stream characteristics requiring design consideration. The baghouse and associated ductwork should be insulated and possibly heated if condensation may occur. Both the structural and fabric components must be considered, as either may be damaged. Where structural corrosion is likely, stainless steel substitution for mild steel may be required, provided that chlorides are not present. (Most austenitic stainless steels are susceptible to chloride corrosion.)

5.2.4.1 Temperature

The temperature of the pollutant stream to be cleaned must be above and remain above the dew point of any condensables in the stream. If the temperature is high and it can be lowered without approaching the dew point, spray coolers or dilution air can be used to drop the temperature so that temperature limits of the fabric will not be exceeded. The additional cost of a precooler will have to be weighed against the higher cost of bags with greater temperature resistance. The use of dilution air to cool the stream also constitutes a tradeoff between a less expensive fabric and a larger filter necessary to accommodate the additional volume of the dilution air. Generally, precooling would not be necessary if fabric that will handle the temperature and the chemical action of the pollutant stream is available. (Costs for spray chambers, quenchers, and other precoolers are found in the "Precoolers" section of the *Manual.*) Table 5.5 lists several of the fabrics in current use and provides information on temperature limits and chemical resistance. The column labeled "Flex Abrasion" indicates the fabric's suitability for cleaning by mechanical shakers.

5.2.4.2 Pressure

Standard fabric filters can be used in pressure or vacuum service but only within the range of about ±25 inches of water. Because of the sheet metal construction of the house, they are not generally suited for more severe service. However, for special applications, high-pressure shells can be built.

Table 5.5: Properties of Leading Fabric Materials[*]

Fabric	Temp, °F[a]	Acid Resistance	Alkali Resistance	Flex Abrasion
Cotton	180	Poor	Very good	Very good
Creslan[b]	250	Good in mineral acids	Good in weak alkali	Good to very good
Dacron[c]	275	Good in most mineral acids; dissolves partially in concentrated H_2SO_4	Good in weak alkali; fair in strong alkali	Very good
Dynel[c]	160	Little effect even at high concentration	Little effect even in high concentration	Fair to good
Fiberglas[d]	500	Fair to good	Fair to good	Fair
Filtron[e]	270	Good to excellent	Good	Good to very good
Gore-Tex[f]	Depends on backing	Depends on backing	Depends on backing	Fair
Nomex[c]	375	Fair	Excellent at low temperature	Excellent
Nylon[c]	200	Fair	Excellent	Excellent
Orlon[c]	260	Good to excellent in mineral acids	Fair to good in weak alkali	Good
Polypropylene	200	Excellent	Excellent	Excellent
Teflon[c]	450	Inert except to fluorine	Inert except to trifluoride, chlorine, and molten alkaline metals	Fair
Wool	200	Very good	Poor	Fair to good

[*] Reference [20]

[a] Maximum continuous operating temperatures recommended by the Industrial Gas Cleaning Institute.

[b] American Cyanamid registered trademark.

[c] Du Pont registered trademark.

[d] Owens-Corning Fiberglas registered trademark.

[e] W. W. Criswell Div. of Wheelabrator-Fry, Inc., trade name.

[f] W. L. Gore and Co., registered trademark.

5.2.5 Equipment Design Considerations

5.2.5.1 Pressure or Suction Housings

The location of the baghouse with respect to the fan in the gas stream affects the capital cost. A suction-type baghouse, with the fan located on the downstream side of the unit, must withstand high negative pressures and therefore must be more heavily constructed and reinforced than a baghouse located downstream of the fan (pressure baghouse). The negative pressure in the suction baghouse can result in outside air infiltration, which can result in condensation, corrosion, or even explosions if combustible gases are being handled. In the case of toxic gases, this inward leakage can have an advantage over the pressure-type baghouse, where leakage is outward. The main advantage of the suction baghouse is that the fan handling the process stream is located at the clean-gas side of the baghouse. This reduces the wear and abrasion on the fan and permits the use of more efficient fans (backward-curved blade design). However, because for some designs the exhaust gases from each compartment are combined in the outlet manifold to the fan, locating compartments with leaking bags may be difficult and adds to maintenance costs. Pressure-type baghouses are generally less expensive because the housing must only withstand the differential pressure across the fabric. In some designs the baghouse has no external housing. Maintenance also is reduced because the compartments can be entered and leaking bags can be observed while the compartment is in service. With a pressure baghouse, the housing acts as the stack to contain the fumes with the subsequent discharge at the roof of the structure, which makes it easier to locate leaking bags. The main disadvantage of the pressure-type baghouse is that the fan is exposed to the dirty gases where abrasion and wear on the fan blades may become a problem.

5.2.5.2 Standard or Custom Construction

The design and construction of baghouses are separated into two groups, standard and custom,[14] which are further separated into low, medium, and high capacity. Standard baghouses are predesigned and factory built as complete off-the-shelf units that are shop-assembled and bagged for low-capacity units (hundreds to thousands of acfm throughput). Medium-capacity units (thousands to less than 100,000 acfm) have standard designs, are shop-assembled, may or may not be bagged, and have separate bag compartment and hopper sections. High-capacity baghouses (larger than 50,000 or 100,000 acfm) can be designed as shippable modules requiring only moderate field assembly. These modules may have bags installed and can be shipped by truck or rail. Upon arrival, they can be operated singly or combined to form units for larger-capacity applications. Because they are preassembled, field labor for installation is less costly.

The custom baghouse, also high capacity, is designed for a specific application and is usually built to the specifications prescribed by the customer. Generally, these units are much larger than standard baghouses. For example, many are used on power plants. The cost of the custom baghouse is much higher per square foot of fabric because it is not an off-the-shelf item and requires special setups for manufacture and expensive field labor for assembly upon arrival. The advantages of the custom baghouse are many and are usually directed towards ease of maintenance, accessibility, and other customer preferences. In some standard baghouses, a complete set of bags must be replaced in a compartment at one time because of the difficulty in locating and replacing single leaking bags, whereas in custom baghouses, single bags are accessible and can be replaced one at a time as leaks develop.

5.2.5.3 Filter Media

The type of filter material used in baghouses is dependent on the specific application in terms of chemical composition of the gas, operating temperature, dust loading, and the physical and chemical characteristics of the particulate. A variety of fabrics, either felted or woven, is available and the selection of a specific material, weave, finish, or weight is based primarily on past experience. The type of yarn (filament, spun, or staple), the yarn diameter, and twist are also factors in the selection of suitable fabrics for a specific application. For some difficult applications, Gore-Tex, a polytetrafluoroethylene (PTFE) membrane laminated to a fabric (felt or woven) may be used. Because of the violent agitation of mechanical shakers, spun or heavy weight staple yarn fabrics are commonly used with this type of cleaning, while lighter weight filament yarn fabrics are used with reverse-air cleaning.

The type of material will limit the maximum operating gas temperature for the baghouse. Cotton fabric has the least resistance to high temperatures (about 180°F), while fiberglass has the most (about 500°F). The temperature of the exhaust-gas stream must be well above the dew point of any of its contained condensables, as liquid particles will usually plug the fabric pores quickly. However, the temperature must be below the maximum limit of the fabric in the bags. These maximum limits are given in Table 5.5.

5.3 Estimating Total Capital Investment

Total capital investment includes costs for the baghouse structure, the initial complement of bags, auxiliary equipment, and the usual direct and indirect costs associated with installing or erecting new structures. These costs are described below. (Costs for improving baghouse performance with electrical enhancement are not discussed in this section, but are mentioned in the example problem.)

5.3.1 Equipment Cost

5.3.1.1 Bare Baghouse Costs

Cost correlations for six types of baghouses are presented. These six types, five of which are preassembled and one, field-assembled, are outlined in Table 5.6.

Each figure gives costs for the filter without bags and additional costs for stainless steel construction and for insulation. All curves are based on a number of actual quotes. A least squares line has been fitted to the quotes and the line's equation is given. In most cases these lines should not be extrapolated in either direction. The reader should not be surprised if he obtains quotes that differ from these curves by as much as ±25%. Significant savings can be obtained by soliciting multiple quotes. All units include inlet and exhaust manifolds, supports, platforms, handrails, and hopper discharge devices. The indicated prices are flange to flange. The scales on both axes change from one figure to another to accommodate the

Table 5.6: Scope of Cost Correlations

Baghouse Type		Figure No.
	Preassembled units	
Intermittent	Shaker	5.2
Continuous	Shaker	5.3
Continuous	Pulse-jet (common housing)	5.4
Continuous	Pulse-jet (modular)	5.5
Continuous	Reverse-air	5.6
	Field-assembled units	
Continuous	Any method	5.7

differing gas flow ranges over which the various types of baghouses operate.

The 304 stainless steel add-on cost is used when such construction is necessary to prevent the exhaust gas stream from corroding the interior of the baghouse. Stainless steel is substituted for all metal surfaces that are in contact with the exhaust gas stream.

Insulation costs are for 3 inches of shop-installed glass fiber encased in a metal skin. One exception is the custom baghouse, which has field-installed insulation. Costs for insulation include only the flange-to-flange baghouse structure on the outside of all areas in contact with the exhaust gas stream. Insulation for ductwork, fan casings, and stacks must be calculated separately as discussed later.

The first baghouse type is the intermittent service baghouse cleaned by a mechanical shaker. This baghouse is shut down and cleaned at convenient times, such as the end of the shift or end of the day. Although few units are sold, they are applicable for operations that require infrequent cleaning. Figure 5.2 presents the unit cost with price in dollars plotted against the gross square feet of cloth required. [21] Because intermittent service baghouses do not require an extra compartment for cleaning, gross and net fabric areas are the same. The plot is linear because baghouses are made up of modular compartments and thus have little economy of

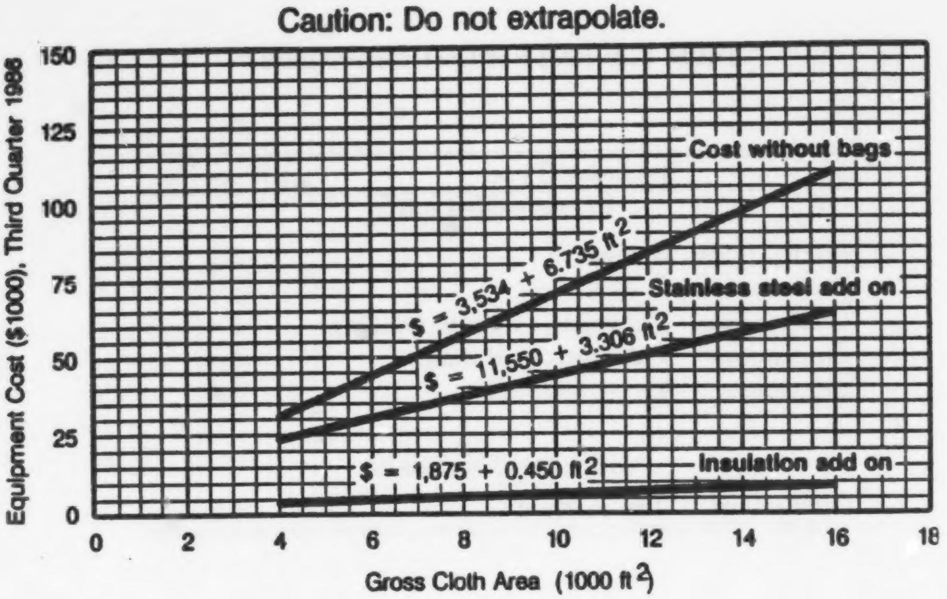
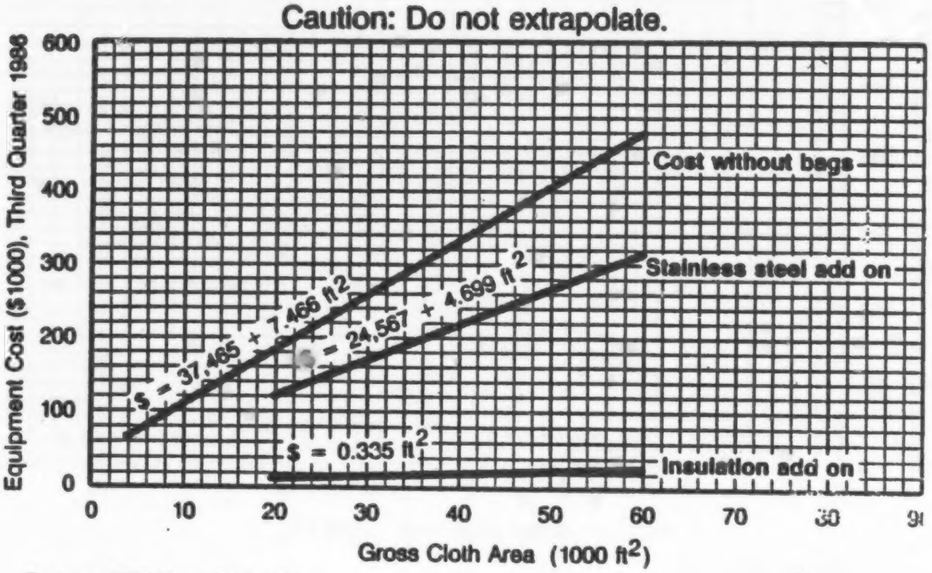

Caution: Do not extrapolate.

Figure 5.2: Equipment costs for intermittent shaker filters

scale. Because of the modular construction, the price line should not be extrapolated downward.

Figure 5.3 presents similar costs for a continuously operated baghouse cleaned by mechanical shaker.[21,22] Again, price is plotted against the gross cloth area in square feet. As in Figure 5.2, the units are modular in construction. Costs for these units, on a square foot basis, are higher because of increased complexity and generally heavier construction.

The third and fourth types are common-housing pulse jets and modular pulse jets. The latter are constructed of separate modules that may be arranged for off-line cleaning, and the former have all bags within one housing. The costs for these units are shown in Figures 5.4 and 5.5, respectively.[21] Note that in the single-unit (common-housing) pulse jet, for the range shown, the height and width of the unit are constant and the length increases; thus, for a different reason than that for the modular units discussed above, the cost increases linearly with size. Because the common



Source: ETS, Inc.; Fuller Co.

Figure 5.3: Equipment costs for continuous shaker filters.

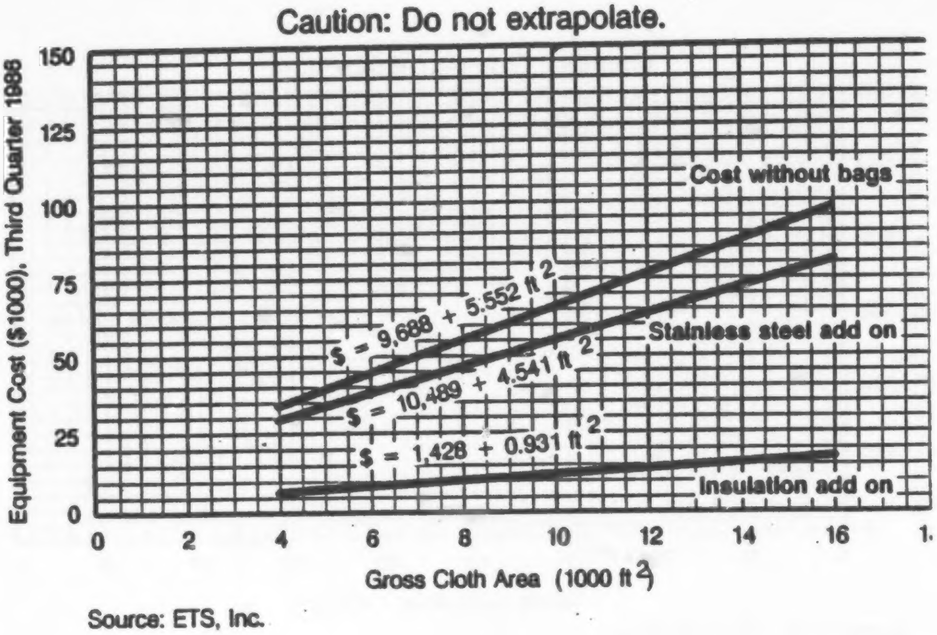
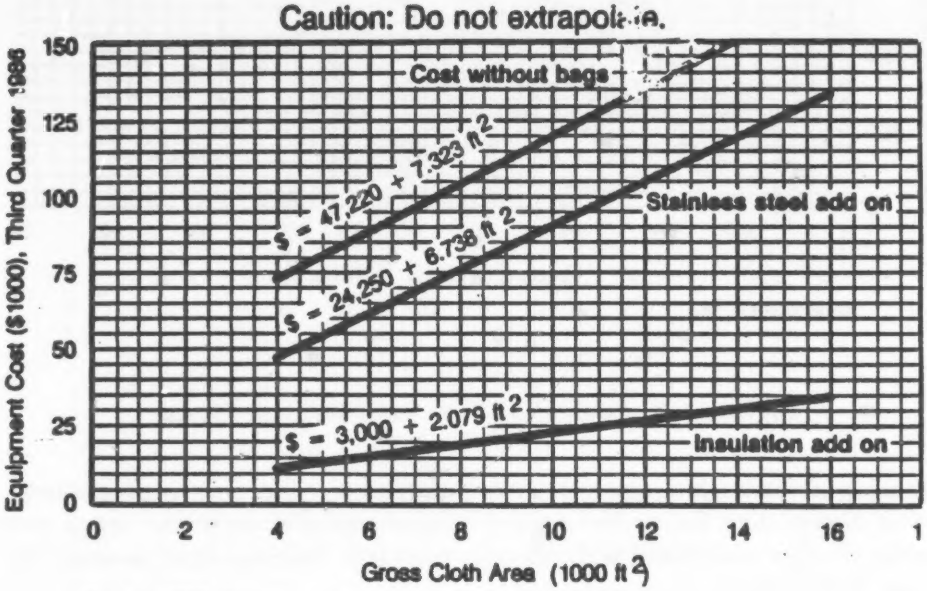
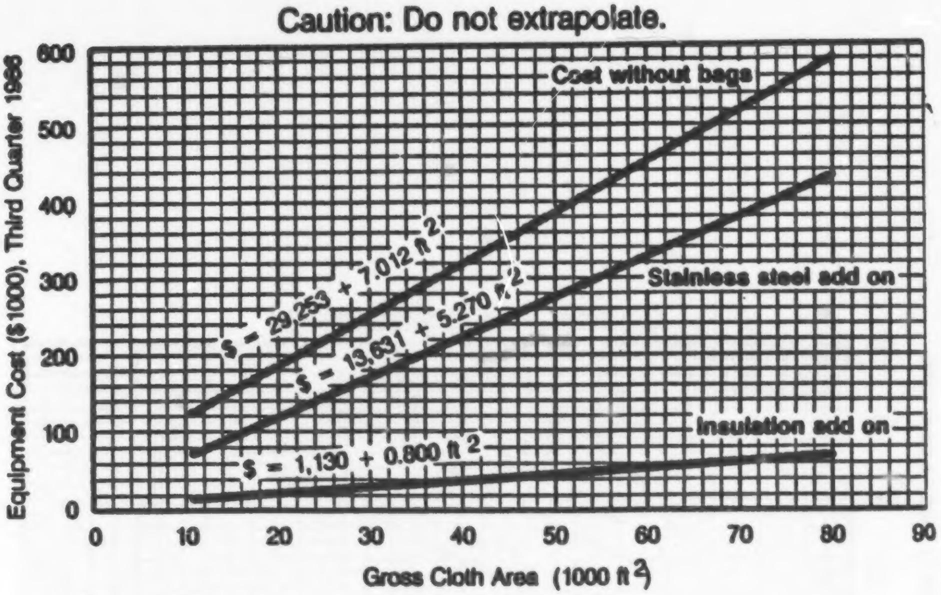


Figure 5.4: Equipment costs for pulse-jet filters (common housing).



Source: ETS, Inc.

Figure 5.5: Equipment costs for pulse-jet filters (modular).



Source: ETS, Inc.

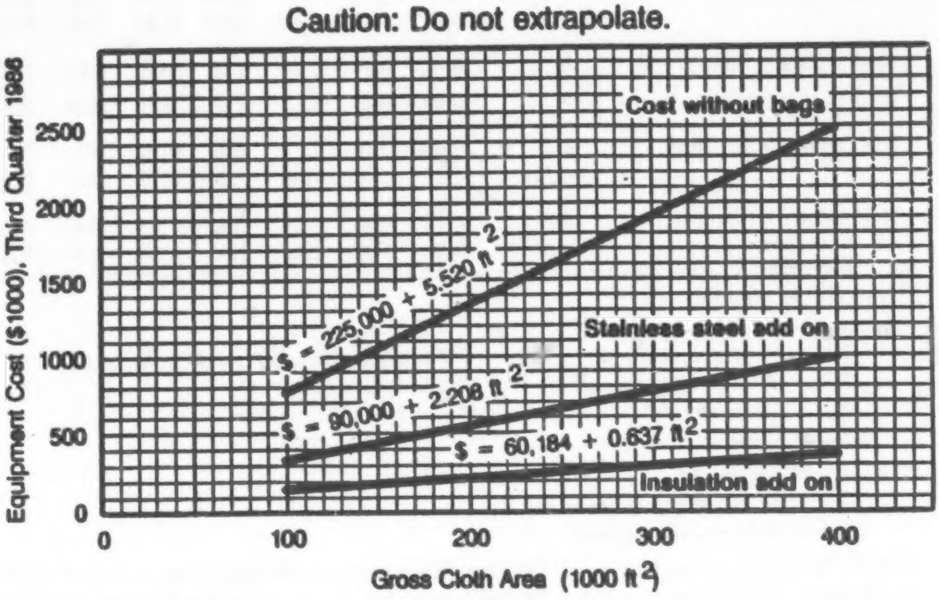
Figure 5.6: Equipment costs for reverse-air filters.

housing is relatively inexpensive, the stainless steel add-on is proportionately higher than for modular units. Added material costs and setup and labor charges associated with the less workable stainless steel account for most of the added expense.

Figure 5.6 shows the costs for the reverse-air baghouses.[21] Again, the construction is modular. The final type is the custom baghouse which, because of its large size, must be field assembled. It is often used on power plants, steel mills, or other applications too large for the factory-assembled baghouses. Prices for custom units are shown in Figure 5.7. [21]

5.3.1.2 Bag Costs

Table 5.7 gives the price per square foot of bags by type of fabric and by type of cleaning system used. Actual quoted prices may vary by ±10% from the values in the table. In calculating the cost, the gross area as



Source: ETS, Inc.

Figure 5.7: Equipment costs for custom-built filters.

Table 5.7: Bag Prices

(3rd quarter 1986 \$/ft^2)

Type of Cleaning	Bag Diameter (inches)	Type of Material[a]						
		PE	PP	NO	HA	FG	CO	TF
Pulse jet, TR[b]	4-1/2 to 5-1/8	0.59	0.61	1.88	0.92	1.29	NA	9.05
	6 to 8	0.43	0.44	1.56	0.71	1.08	NA	6.80
Pulse jet, BBR	4-1/2 to 5-1/8	0.37	0.40	1.37	0.66	1.24	NA	8.78
	6 to 8	0.32	0.33	1.18	0.58	0.95	NA	6.71
Shaker, Strap top	5	0.45	0.48	1.28	0.75	NA	0.44	NA
Shaker, Loop top	5	0.43	0.45	1.17	0.66	NA	0.39	NA
Reverse air with rings	8	0.46	NA	1.72	NA	0.99	NA	NA
	11-1/2	0.47	NA	1.69	NA	0.76	NA	NA
Reverse air w/o rings	8	0.32	NA	1.20	NA	0.69	NA	NA
	11-1/2	0.32	NA	1.16	NA	0.53	NA	NA

NA = Not applicable.

[a] Materials:

PE = 16-os polyester
PP = 16-os polypropylene
NO = 14-os nomex
HA = 16-os homopolymer acrylic
FG = 16-os fiberglass with 10% Teflon
CO = 9-os cotton
TF = 22-os Teflon felt

[b] Bag removal methods:

TR = Top bag removal (snap in)
BBR = Bottom bag removal

NOTE: For pulse-jet baghouses, all bags are felts except for the fiberglass, which is woven. For bottom access pulse jets, the mild steel cage price for one cage can be calculated from the single-bag fabric area using:

$$\$ = 4.941 + 0.163\ \text{ft}^2 \text{ in 50 cage lots}$$
$$\$ = 4.441 + 0.163\ \text{ft}^2 \text{ in 100 cage lots}$$
$$\$ = 3.941 + 0.163\ \text{ft}^2 \text{ in 500 cage lots}$$

These costs apply to 4-1/2 inch or 5-5/8 inch diameter, 8 foot and 10 foot cages made of 11 gauge mild steel and having 10 vertical wires and "Roll Band" tops. For flanged tops, add \$1 per cage. If flow control venturis are used (as they are in about half of the pulse-jet manufacturers' designs), add \$5 per cage. For stainless steel cages use:

$$\$ = 23.335 + 0.280 \text{ ft2 in 50 cage lots}$$
$$\$ = 21.791 + 0.263 \text{ ft2 in 100 cages lots}$$
$$\$ = 20.564 + 0.248 \text{ ft2 in 500 cage lots}$$

For shakers and reverse air baghouses, all bags are woven. All prices are for finished bags, and prices can vary from one supplier to another. For Gore-Tex bag prices, multiply base fabric price by factors of 3 to 4.5.

Source: ETS, Inc.[21]

determined from Table 5.2 should be used. Gore-Tex fabric costs are a combination of the base fabric cost and a premium for the PTFE laminate and its application. As fiber market conditions change, the costs of fabrics relative to each other also change. The bag prices are based on typical fabric weights, in ounces/square yard, for the fabric being priced. Sewn-in snap rings are included in the price, but other mounting hardware, such as clamps or cages, is an added cost.

5.3.1.3 Auxiliary Equipment

The auxiliary equipment depicted in Figure 5.1 is discussed elsewhere in the *Manual*. Because hoods, ductwork, precoolers, cyclones, fans, motors, dust removal equipment and stacks are common to many pollution control systems, they are (or will be) given extended treatment in separate chapters.

5.3.2 Total Purchased Cost

The total purchased cost of the fabric filter system is the sum of the costs of the baghouse, bags, auxiliary equipment, instruments and controls; and of taxes and freight. The last three items generally are taken as percentages of the estimated total cost of the first three items. Typical values, from Chapter 2, are 10% for instruments and controls, 3% for taxes, and 5% for freight.

Bag costs can vary from less than 15% to more than 100% of bare baghouse cost, depending on type of fabric required. This situation makes it inadvisable to estimate total purchased cost without considering both costs, and prevents effective use of factors to estimate a single cost for the baghouse and bags.

5.3.3 Total Capital Investment

Using Chapter 2 methodology, the total capital investment (TCI) for most baghouses is estimated from a series of factors applied to the purchased equipment cost to obtain direct and indirect installation costs. The TCI is the sum of these three costs (*i.e.*, purchased equipment and direct and indirect installation costs). The required factors are given in Table 5.8. Because bag costs can have such a large effect on the total purchased equipment cost, the factors may cause overestimation of total capital investment when expensive bags are used. Using stainless steel components may also cause overestimation. Because baghouses may vary from small units installed within existing buildings to large, separate structures, specific factors for site preparation or for buildings are not given. However, costs for buildings may be obtained from such references as Means Square Foot Costs 1986.[23] Land, working capital, and off-site facilities are excluded from the table, as they are not normally required. For very large installations, however, they may be needed and would be estimated on an as-needed basis.

The factors given in Table 5.8 are for average installation conditions. Considerable variation may be seen with other-than-average installation circumstances. Moreover, the Table 5.8 factors may be too large for "packaged" fabric filters—those pre-assembled baghouses that consist of the compartments, bags, waste gas fan and motor, and instruments and controls. Because these packaged units require very little installation, their installation costs would be lower (20-25% of the purchased equipment cost).

5.4 Estimating Total Annual Costs

5.4.1 Direct Annual Cost

Direct annual costs include operating and supervisory labor, operating materials, replacement bags, maintenance (labor and materials), utilities, and dust disposal. Most of these costs are discussed individually below. They vary considerably with location and time, and, for this reason, should be obtained to suit the specific baghouse system being costed. For example, current labor rates may be found in such publications as the *Monthly Labor Review*, published by the U.S. Department of Labor, Bureau of Labor Statistics.

Table 5.8: Capital Cost Factors for Fabric Filters[a]

Cost Item	Factor
Direct costs	
Purchased equipment costs	
Fabric filter (EC) + bags + auxiliary equipment	As estimated, A
Instrumentation	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased Equipment Cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.04 B
Handling & erection	0.50 B
Electrical	0.08 B
Piping	0.01 B
Insulation for ductwork[b]	0.07 B
Painting[c]	0.02 B
Direct installation cost	0.72 B
Site preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Cost	1.72 B + SP + Bldg.
Indirect Costs (installation)	
Engineering	0.10 B
Construction and field expense	0.20 B
Contractor fees	0.10 B
Start-up	0.01 B
Performance test	0.01 B
Contingencies	0.03 B
Total Indirect Cost, IC	0.45 B
Total Capital Investment = DC + IC	2.17 B + SP + Bldg.

[a] Reference [24]

[b] If ductwork dimensions have been established, cost may be estimated based on \$10 to \$12/ft^2 (fourth quarter 1986) of surface for field application. Fan housings and stacks may also be insulated.[21]

[c] The increased use of special coatings may increase this factor to 0.06B or higher.[25]

5.4.1.1 Operating and Supervisory Labor

Typical operating labor requirements are 2 to 4 hours per shift for a wide range of filter sizes.[24] Small or well-performing units may require less time, while very large or troublesome units may require more. Supervisory labor is taken as 15% of operating labor.

5.4.1.2 Operating Materials

Operating materials are generally not required for baghouses. An exception is the use of precoat materials injected on the inlet side of the baghouse to provide a protective dust layer on the bags when sticky or corrosive particles might harm them. Adsorbents may be similarly injected when the baghouse is used for simultaneous particle and gas removal. Costs for these materials should be included on a dollars-per-mass basis (*e.g.*, dollars per ton).

5.4.1.3 Maintenance

Maintenance labor varies from 1 to 2 hours per shift.[24] As with operating labor, these values may be reduced or exceeded depending on the size and operating difficulty of a particular unit. Maintenance materials costs are assumed to be equal to maintenance labor costs.[24]

5.4.1.4 Replacement Parts

The major replacement part items are filter bags, which have a typical operating life of about 2 years. The following formula is used for computing the bag replacement cost:

$$CRC_B = (C_B + C_L) \times CRF_B \tag{5.13}$$

where

CRC_B = bag capital recovery cost (\$/year)

C_B = initial bag cost including taxes and freight (\$)

C_L = bag replacement labor (\$)

CRF_B = capital recovery factor whose value is a function of the annual interest rate and the useful life of the bags (For instance, for a 10% interest rate and a 2-year life, $CRF_B = 0.5762$.)

The bag replacement labor cost (C_L) will depend on such factors as the number, size, and type of the bags; their accessibility; how they are connected to the baghouse tube-sheet; etc. For example, in a reverse-air baghouse it would probably take from 10 to 20 man-minutes to change an 8 inch by 24 foot bag that is clamped in place. This bag has a filtering surface area of approximately 50 ft^2. If the replacement labor rate were \$21.12/h (including overhead), C_L would be from \$0.07 to \$0.14/ft^2 of bag area. As Table 5.7 shows, for some bags (*e.g.*, cotton), this range of C_L would constitute a significant fraction of the purchased cost. For pulse jets, replacement time would be about 5 to 10 man-minutes for a 5 inch by 10 foot bag in a top-access baghouse. These bag replacement times are based on changing a minimum of an entire module and on having typical baghouse designs. Times would be significantly longer if only a few bags were being replaced or if the design for bag attachment or access were atypical.

This method treats the bags as an investment that is amortized over the useful life of the bags, while the rest of the control system is amortized over its useful life (typically 20 years; see Subsection 5.4.2). Values of CRF_B for bag lives different from 2 years can be calculated from Equation 2.3.

5.4.1.5 Electricity

Power is required to operate system fans and cleaning equipment. Fan power for primary gas movement can be calculated from Equation 2.7. After substituting into this equation a combined fan-motor efficiency of 0.65 and a specific gravity of 1.000, we obtain:[26]

$$\text{Power}_{fan} = 0.000181 Q (\Delta P) \theta \tag{5.14}$$

where

Power_{fan} = fan power requirement (kWh/yr)
Q = system flow rate (acfm)
ΔP = system pressure drop (in. H_2O)
θ = operating time (h/yr)

Cleaning energy for reverse-air systems can be calculated from the number of compartments to be cleaned at one time (usually one, sometimes two), and the reverse gas-to-cloth ratio (from about one to two times the forward gas-to-cloth ratio). Reverse-air pressure drop varies up to 6 or 7 in. H_2O depending on location of the fan pickup (before or after the main system fan).[27] The reverse-air fan generally runs continuously.

Typical energy consumption in kWh/yr for a shaker system operated 8,760 h/yr can be calculated from:[5]

$$P = 0.053A \tag{5.15}$$

where

A = gross fabric area (ft^2)

5.4.1.6 Fuel

If the baghouse or associated ductwork is heated to prevent condensation, fuel costs should be calculated as required. These costs can be significant, but may be difficult to predict. For methods of calculating heat transfer requirements, see Perry.[26]

5.4.1.7 Water

Cooling process gases to acceptable temperatures for fabrics being used can be done by dilution with air, evaporation with water, or heat exchange with normal equipment. The last two cases require consumption of plant water, although costs are not usually significant. Section 4.4 of Chapter 4 provides information on estimating cooling-water costs.

5.4.1.8 Compressed Air

Pulse-jet filters use compressed air at pressures of about 60 to 100 psig. Typical consumption is about 2 scfm/1,000 cfm of gas filtered.[5] For example, a unit filtering 20,000 cfm of gas uses about 40 scfm of compressed air for each minute the filter is operated.

5.4.1.9 Dust Disposal

If collected dust cannot be recycled or sold, it must be landfilled or disposed of in some other manner. Disposal costs are site-specific, but they may typically run $20 or $30 per ton exclusive of transportation (see Section 2.4, Chapter 2).

5.4.2 Indirect Annual Cost

These include such costs as capital recovery, property taxes, insurance, administrative costs ("G&A"), and overhead. The capital recovery cost is based on the equipment lifetime and the annual interest rate employed. (See Chapter 2 for a thorough discussion of the capital recovery cost and the variables that determine it.) For fabric filters, the system lifetime varies from 5 to 40 years, with 20 years being typical.[24] However, this does not apply to the bags, which usually have much shorter lives. (See Section 5.4.1.) Therefore, as Chapter 2 suggests, when figuring the system capital recovery cost, one should base it on the installed capital cost less the cost of replacing the bags (*i.e.*, the purchased cost of the bags plus the cost of labor necessary to replace them). In other words:

$$CRC_s = [TCI - C_B - C_L]CRF_s \tag{5.16}$$

where

CRC_s = capital recovery cost for fabric filter system ($/yr)
TCI = total capital investment ($)
C_B = initial cost of bags including taxes and freight ($)
C_L = labor cost for replacing bags ($)
CRF_s = capital recovery factor for fabric filter system (defined in Chapter 2).

For example, for a 20-year system life and a 10% annual interest rate, the CRF, would be 0.1175.

As Chapter 2 suggests, the suggested factor to use for property taxes, insurance, and administrative charges is 4% of the TCI. Finally, the overhead is calculated as 60% of the sum of operating, supervisory, and maintenance labor, and maintenance materials.

5.4.3 Recovery Credits

For processes that can reuse the dust collected in the baghouse or that can sell the dust in a local market, such as fly ash sold as an extender for paving mixes, a credit should be taken. As used below, this credit (RC) appears as a negative cost.

5.4.4 Total Annual Cost

Total annual cost for owning and operating a fabric filter system is the sum of the components listed in Sections 5.4.1 through 5.4.3, *i.e.*:

$$TAC = DC + IC - RC \tag{5.17}$$

where

TAC = total annual cost ($)
DC = direct annual cost ($)
IC = indirect annual cost ($)
RC = recovery credits (annual) ($)

5.4.5 Example Problem

Assume a baghouse is required for controlling fly ash emissions from a coal-fired boiler. The flue gas stream is 50,000 acfm at 325°F and has an ash loading of 4 gr/ft^3. Analysis of the ash shows a mass median diameter of 7 μm. Assume the baghouse operates for 8,640 h/yr (360 d).

The gas-to-cloth ratio (G/C) can be taken from Table 5.1 as 2.5, for woven fabrics in shaker or reverse-air baghouses, or 5, for felts used in pulse-jet baghouses. If a factor method were used for estimating G/C, Table 5.3 for shakers would yield the following values: A = 2, B = 0.9, and C = 1.0. The gas-to-cloth ratio would be:

$$2 \times 0.9 \times 1.0 = 1.8.$$

This value could also be used for reverse-air cleaning. For a pulse-jet unit, Table 5.4 gives a value of 9.0 for factor A and 0.8 for factor B. Equation 5.11 becomes:

$$\begin{aligned} V &= 2.878 \times 9.0 \times 0.8(275)^{-0.2335}(4)^{-0.06021}(0.7471 + 0.0853 \ln 7) \quad (5.18) \\ &= 4.69 \end{aligned}$$

Because this value is so much greater than the shaker/reverse-air G/C, we conclude that the pulse-jet baghouse would be the least costly design.[2] Assume the use of on-line cleaning in a common housing structure and, due to the high operating temperature, the use of glass filter bags (see Table 5.5). At a gas-to-cloth ratio of 4.69, the fabric required is[3]

$$50,000 \text{ acfm}/4.69 \text{ fpm} = 10,661 \text{ ft}^2.$$

From Figure 5.4, the cost of the baghouse ("common housing" design) is:

$$\text{Cost} = 9,688 + 5.552(10,661) = \$68,878 \quad (5.19)$$

Insulation is required. The insulation add-on cost from Figure 5.4 is:

$$\text{Cost} = 1,428 + 0.931(10,661) = \$11,353 \quad (5.20)$$

From Table 5.7, bag costs are \$1.24/ft^2 for 5-1/8 inch diameter glass fiber, bottom removal bags. Total bag cost is

$$10,661 \text{ ft}^2 \times \$1.24/\text{ft}^2 = \$13,220.$$

For 10 ft long cages,

---

[2] This conclusion is based on the inference that a much higher G/C would yield lower capital and, in turn, annual costs. However, to make a more rigorous selection, we would need to calculate and compare the total annual costs of all three baghouse designs (assuming all three are technically acceptable). The reader is invited to make this comparison. Further discussion of the effects of G/C increases, and accompanying pressure drop increases, on overall annual costs will be found in Reference 27.

[3] This is the total (gross) bag area required. No bag area adjustment factor has been applied here, because this is a common housing pulse jet unit that is cleaned continuously during operation. Thus, no extra bag compartment is needed, and the gross and net bag areas are equal.

$$\begin{aligned} \text{fabric area per cage} &= 5\tfrac{1}{8}\text{ in./12 in./ft} \times \pi \times 10\text{ ft} \\ &= 13.42\text{ ft}^2. \\ \text{The number of cages} &= 10{,}661\text{ ft}^2/13.42\text{ ft}^2 \\ &= 795\text{ cages (rounded up to next integer).} \end{aligned}$$

From Table 5.7, individual cage cost is

$$3.941 + 0.163(13.42\text{ft}^2) = \$6.128.$$

Total cage cost is

$$795\text{ cages} \times \$6.128/\text{cage} = \$4,872.$$

Assume the following auxiliary costs have been estimated from data in other parts of the *Manual*:

Ductwork	$14,000
Fan	14,000
Motor	7,000
Starter	3,500
Dampers	7,200
Compressor	6,000
Screw conveyor	4,000
Stack	7,000
Total	$62,700

Direct costs for the fabric filter system, based on the factors in Table 5.8, are given in Table 5.9. (Again, we assume site preparation and buildings costs to be negligible.) Total capital investment is $412,000. Table 5.10 gives the direct and indirect annual costs, as calculated from the factors given in Section 5.4.1. For bag replacement labor, assume 10 min per bag for each of the 795 bags. At a maintenance labor rate of $21.12 (including overhead), the labor cost is $2,809 for 133 h. The bags are assumed to be replaced every 2 yr. The replacement cost is calculated using Equation 5.13.

Pressure drop (for energy costs) can be calculated from Equations 5.8 and 5.9, with the following assumed values:

$$K_2 = 15\frac{\text{in. H}_2\text{O}/(\text{ft/min})}{\text{lb/ft}^2}$$

Table 5.9: Capital Costs for Fabric Filter System
Example Problem

Cost Item	Cost
Direct Costs	
Purchased equipment costs	
Fabric filter (with insulation)(EC)	$80,231
Bags and cages	18,092
Auxiliary equipment	62,700
Sum = A	$161,023
Instrumentation, 0.1A	16,102
Sales taxes, 0.03A	4,831
Freight, 0.05A	8,051
Purchased equipment cost, B	$190,007
Direct installation costs	
Foundation and supports, 0.04B	7,600
Handling and erection, 0.50B	95,004
Electrical, 0.08B	15,201
Piping, 0.01B	1,900
Insulation for ductwork, 0.07B	13,300
Painting, 0.02B	3,800
Direct installation cost	$136,805
Site preparation	—
Facilities and buildings	—
Total Direct Cost	$326,812
Indirect Costs (installation)	
Engineering, 0.10B	19,001
Construction and field expenses, 0.20B	38,001
Contractor fees, 0.10B	19,001
Start-up, 0.01B	1,900
Performance test, 0.01B	1,900
Contingencies, 0.03B	5,700
Total Indirect Cost	$85,503
Total Capital Investment (rounded)	$412,000

Table 5.10: Annual Costs for Fabric Filter System Example Problem

Cost Item	Calculations	Cost
Direct Annual Costs, DC		
Operating labor		
Operator	$\frac{6\text{ h}}{\text{day}} \times \frac{360\text{ days}}{\text{yr}} \times \frac{\$12}{\text{h}}$	$25,920
Supervisor	15% of operator = .15 × 25,920	3,888
Operating materials		—
Maintenance		
Labor	$\frac{3\text{ h}}{\text{day}} \times \frac{360\text{ days}}{\text{yr}} \times \frac{\$13.20}{\text{h}}$	14,256
Material	100% of maintenance labor	14,256
Replacement parts, bags	$[2,809 + (13,220 \times 1.08^{a})] \times 0.5762$	9,845
Utilities		
Electricity	$0.000181 \times 50,000\text{ acfm} \times 10.3\text{ in. H}_2\text{O} \times \frac{8,640\text{ h}}{\text{yr}} \times \frac{\$0.06}{\text{kWh}}$	48,323
Compressed air (dried and filtered)	$\frac{2\text{ scfm}}{1,000\text{ acfm}} \times 50,000\text{ acfm} \times \frac{\$0.16}{1,000\text{ scfm}} \times \frac{60\text{ min}}{\text{h}} \times \frac{8,640\text{ h}}{\text{yr}}$	8,294
Waste disposal	at $20/ton on-site for essentially 100% collection efficiency: $\frac{4\text{ gr}}{\text{ft}^3} \times \frac{1\text{ lb}}{7,000\text{ gr}} \times \frac{50,000\text{ ft}^3}{\text{min}} \times \frac{60\text{ min}}{\text{h}} \times \frac{8,640\text{ h}}{\text{yr}} \times \frac{1\text{ ton}}{2,000\text{ lb}} \times \frac{\$20}{\text{ton}}$	148,114
Total DC		$272,896
Indirect Annual Costs, IC		
Overhead	60% of sum of operating, supv., & maint. labor & maint. materials = 0.6(25,920 + 3,888 + 14,256 + 14,256) =	34,992
Administrative charges	2% of Total Capital Investment = 0.02($412,315)	8,246
Property tax	1% of Total Capital Investment = 0.01($412,315)	4,123
Insurance	1% of Total Capital Investment = 0.01($412,315)	4,123
Capital recovery[b]	0.1175(412,315 − 2,809 − 13,220 × 1.08)	46,439
Total IC		$97,923
Total Annual Cost (rounded)		$371,000

[a] The 1.08 factor is for freight and sales taxes.

[b] The capital recovery cost factor, CRF, is a function of the fabric filter or equipment life and the opportunity cost of the capital (*i.e.*, interest rate). For example, for a 20 year equipment life and a 10% interest rate, CRF = 0.1175.

$$
\begin{aligned}
P_j &= 100 \text{ psig} \\
\text{cleaning interval} &= 10 \text{ min}
\end{aligned}
$$

We further assume that a G/C of 4.69 ft/min is a good estimate of the mean face velocity over the duration of the filtering cycle.

$$
\begin{aligned}
W_o &= C_i V \theta \\
&= 4 \frac{\text{gr}}{\text{ft}^3} \times \frac{1 \text{ lb}}{7{,}000 \text{ gr}} \times 4.69 \frac{\text{ft}}{\text{min}} \times 10 \text{ min} \\
&= 0.0268 \text{ lb/ft}^2 \\
\Delta P &= 6.08 \times 4.69 \frac{\text{ft}}{\text{min}} \times (100 \text{ psig})^{-0.65} + 15 \frac{\text{in. H}_2\text{O/(ft/min)}}{\text{lb/ft}^2} \\
&\quad \times 0.0268 \frac{\text{lb}}{\text{ft}^2} \times 4.69 \frac{\text{ft}}{\text{min}} \\
&= 3.32 \text{ in. H}_2\text{O across the fabric (when fully loaded).}
\end{aligned}
$$

We will assume that the baghouse structure and the ductwork contribute an additional 3 in. H_2O and 4 in. H_2O, respectively. The total pressure drop is, therefore, 10.3 inches.

The total annual cost is $371,000, nearly half of which is for ash disposal. If a market for the fly ash could be found, the total annual cost would be greatly reduced. For example, if $2/ton were received for the ash, the total annual cost would drop to $208,000 ($371,000 - $148,000 - $14,800), or 56% of the cost when no market exists. Clearly, the total annual cost is extremely sensitive to the value chosen for the dust disposal cost in this case. In this and in similar cases, this value should be selected with care.

As discussed in the Design Procedures Section (Section 5.2), an electrostatically enhanced baghouse (an emerging technology) may have up to 30% lower total annual costs than the baghouse estimated in the example problem.

5.5 Acknowledgment

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Chapter 6

ELECTROSTATIC PRECIPITATORS

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Contents

6.1 Process Description 6-4

- 6.1.1 Introduction 6-4
- 6.1.2 Types of ESPs 6-5
  - 6.1.2.1 Plate-Wire Precipitators 6-5
  - 6.1.2.2 Flat Plate Precipitators 6-10
  - 6.1.2.3 Tubular Precipitators 6-10
  - 6.1.2.4 Wet Precipitators 6-11
  - 6.1.2.5 Two-Stage Precipitators 6-11
- 6.1.3 Auxiliary Equipment 6-12
- 6.1.4 Electrostatic Precipitation Theory 6-14
  - 6.1.4.1 Electrical Operating Point 6-15
  - 6.1.4.2 Particle Charging 6-17
  - 6.1.4.3 Particle Collection 6-19
  - 6.1.4.4 Sneakage and Rapping Reentrainment 6-21

6.2 ESP Design Procedure 6-23

- 6.2.1 Specific Collecting Area 6-23
  - 6.2.1.1 SCA Procedure with Known Migration Velocity 6-24
  - 6.2.1.2 Full SCA Procedure 6-26
  - 6.2.1.3 Specific Collecting Area for Tubular Precipitators 6-34

6.2.2 Flow Velocity 6-34

6.2.3 Pressure Drop Calculations 6-36

6.2.4 Particle Characteristics 6-37

6.2.5 Gas Characteristics 6-39

6.2.6 Cleaning 6-40

6.2.7 Construction Features 6-40

6.3 Estimating Total Capital Investment 6-42

6.3.1 Equipment Cost 6-42

6.3.1.1 ESP Costs 6-42

6.3.1.2 Retrofit Cost Factor 6-46

6.3.1.3 Auxiliary Equipment 6-47

6.3.1.4 Costs for Two-Stage Precipitators 6-47

6.3.2 Total Purchased Cost 6-49

6.3.3 Total Capital Investment (TCI) 6-50

6.4 Estimating Total Annual Costs 6-50

6.4.1 Direct Annual Costs 6-50

6.4.1.1 Operating and Supervisory Labor 6-52

6.4.1.2 Operating Materials 6-53

6.4.1.3 Maintenance 6-53

6.4.1.4 Electricity 6-53

6.4.1.5 Fuel 6-55

6.4.1.6 Water 6-55

6.4.1.7 Compressed Air 6-55
6.4.1.8 Dust Disposal 6-55
6.4.1.9 Wastewater Treatment 6-56
6.4.1.10 Conditioning Costs 6-56
6.4.2 Indirect Annual Costs 6-56
6.4.3 Recovery Credits 6-57
6.4.4 Total Annual Cost 6-57
6.4.5 Example Problem 6-58
6.4.5.1 Design SCA 6-58
6.4.5.2 ESP Cost 6-62
6.4.5.3 Costs of Auxiliaries 6-62
6.4.5.4 Total Capital Investment 6-63
6.4.5.5 Annual Costs–Pressure Drop 6-63
6.4.5.6 Total Annual Cost 6-63
6.5 Acknowledgments 6-66
Appendix 6A – Effects of Material Thickness and Type on ESP Costs 6-67
References 6-73

6.1 Process Description

6.1.1 Introduction

An electrostatic precipitator (ESP) is a particle control device that uses electrical forces to move the particles out of the flowing gas stream and onto collector plates. The particles are given an electrical charge by forcing them to pass through a corona, a region in which gaseous ions flow. The electrical field that forces the charged particles to the walls comes from electrodes maintained at high voltage in the center of the flow lane.

Once the particles are collected on the plates, they must be removed from the plates without reentraining them into the gas stream. This is usually accomplished by knocking them loose from the plates, allowing the collected layer of particles to slide down into a hopper from which they are evacuated. Some precipitators remove the particles by intermittent or continuous washing with water.

Figure 6.1 is an illustration of an ESP with its various components identified. Figure 6.2 shows two variations of charging electrode/collector electrode arrangements used in ESPs.

6.1.2 Types of ESPs

ESPs are configured in several ways. Some of these configurations have been developed for special control action, and others have evolved for economic reasons. The types that will be described here are (1) the plate-wire precipitator, the most common variety; (2) the flat plate precipitator; (3) the tubular precipitator; (4) the wet precipitator, which may have any of the previous mechanical configurations; and (5) the two-stage precipitator.

6.1.2.1 Plate-Wire Precipitators

Plate-wire ESPs are used in a wide variety of industrial applications, including coal-fired boilers, cement kilns, solid waste incinerators, paper mill recovery boilers, petroleum refining catalytic cracking units, sinter plants, basic oxygen furnaces, open hearth furnaces, electric arc furnaces, coke oven batteries, and glass furnaces.

In a plate-wire ESP, gas flows between parallel plates of sheet metal and high-voltage electrodes. These electrodes are long wires weighted and hanging between the plates or are supported there by mast-like structures (rigid frames). Within each flow path, gas flow must pass each wire in sequence as it flows through the unit.

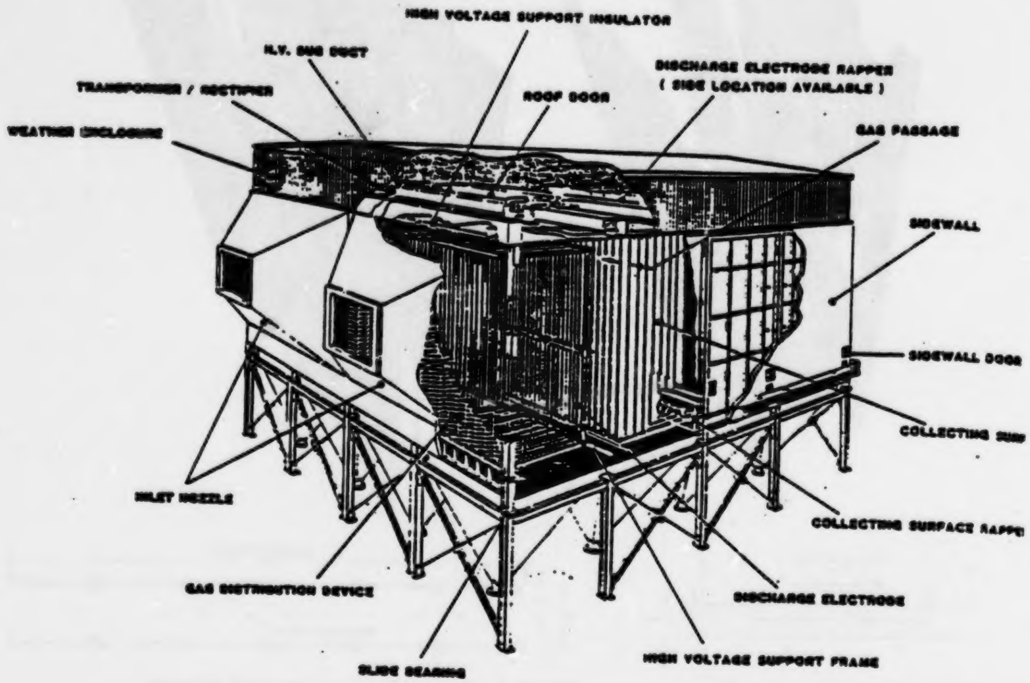
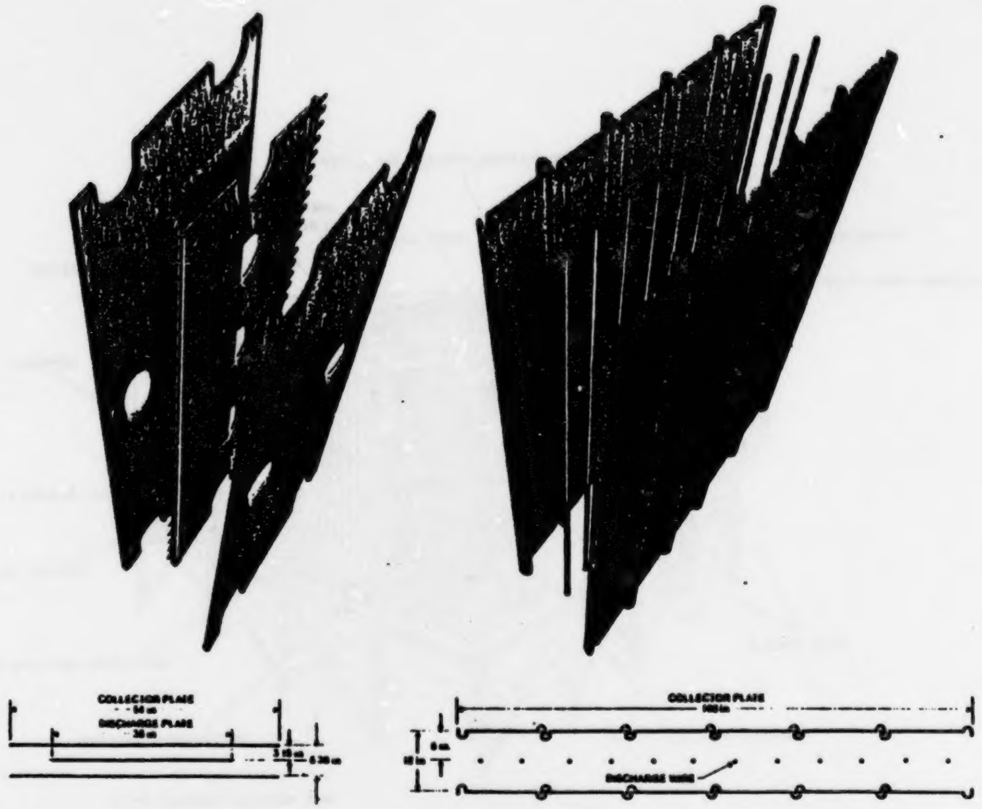


Figure 6.1: Electrostatic Precipitator Components

(Courtesy of the Industrial Gas Cleaning Institute)



Plan view of McGill EP electrodes with typical dimensions

Plan view of Conventional EP electrodes with typical dimensions

Figure 6.2: Flat-plate and Plate-wire ESP Configurations

(Courtesy of United McGill Corporation)

The plate-wire ESP allows many flow lanes to operate in parallel, and each lane can be quite tall. As a result, this type of precipitator is well suited for handling large volumes of gas. The need for rapping the plates to dislodge the collected material has caused the plate to be divided into sections, often three or four in series with one another, which can be rapped independently. The power supplies are often sectionalized in the same way to obtain higher operating voltages, and further electrical sectionalization may be used for increased reliability. Dust also deposits on the discharge electrode wires and must be periodically removed similarly to the collector plate.

The power supplies for the ESP convert the industrial ac voltage (220 to 480 V) to pulsating dc voltage in the range of 20,000 to 100,000 V as needed. The supply consists of a step-up transformer, high-voltage rectifiers, and sometimes filter capacitors. The unit may supply either half-wave or full-wave rectified dc voltage. There are auxiliary components and controls to allow the voltage to be adjusted to the highest level possible without excessive sparking and to protect the supply and electrodes in the event a heavy arc or short-circuit occurs.

The voltage applied to the electrodes causes the air between the electrodes to break down electrically, an action known as a "corona". The electrodes usually are given a negative polarity because a negative corona supports a higher voltage than a positive corona before sparking occurs. The ions generated in the corona follow electric field lines from the wires to the collecting plates. Therefore, each wire establishes a charging zone through which the particles must pass.

Particles passing through the charging zone intercept some of the ions, which become attached. Small aerosol particles (<1 μm diameter) can absorb tens of ions before their total charge becomes large enough to repel further ions, and large particles (>10 μm diameter) can absorb tens of thousands. The electrical forces are therefore much stronger on the large particles.

As the particles pass each successive wire, they are driven closer and closer to the collecting walls. The turbulence in the gas, however, tends to keep them uniformly mixed with the gas. The collection process is therefore a competition between the electrical and dispersive forces. Eventually, the particles approach close enough to the walls so that the turbulence drops to low levels and the particles are collected.

If the collected particles could be dislodged into the hopper without losses, the ESP would be extremely efficient. The rapping that dislodges the accumulated layer also projects some of the particles (typically 12 percent for coal fly ash) back into the gas stream. These reentrained particles are then processed again by later sections, but the particles reentrained in the last section of the ESP have no chance to be recaptured and so escape the unit.

Practical considerations of passing the high voltage into the space between the lanes and allowing for some clearance above the hoppers to support and align electrodes leave room for part of the gas to flow around the charging zones. This is called "sneakage" and amounts to 5 to 10 percent of the total flow. Antisneakage baffles usually are placed to force the sneakage flow to mix with the main gas stream for collection in later sections. But, again, the sneakage flow around the last section has no opportunity to be collected.

These losses play a significant role in the overall performance of an ESP. Another major factor is the resistivity of the collected material. Because the particles form a continuous layer on the ESP plates, all the ion current must pass through the layer to reach the ground plates. This current creates an electric field in the layer, and it can become large enough to cause local electrical breakdown. When this occurs, new ions of the wrong polarity are injected into the wire-plate gap where they reduce the charge on the particles and may cause sparking. This breakdown condition is called "back corona".

Back corona is prevalent when the resistivity of the layer is high, usually above 2×10^{11} ohm-cm. For lower resistivities, the operation of the ESP is not impaired by back corona, but resistivities much higher than 2×10^{11} ohm-cm considerably reduce the collection ability of the unit because the severe back corona causes difficulties in charging the particles. At resistivities below 10^{8} ohm-cm, the particles are held on the plates so loosely that rapping and nonrapping reentrainment become much more severe. Care must be taken in measuring or estimating resistivity because it is strongly affected by variables such as temperature, moisture, gas composition, particle composition, and surface characteristics.

6.1.2.2 Flat Plate Precipitators

A significant number of smaller precipitators (100,000 to 200,000 acfm) use flat plates instead of wires for the high-voltage electrodes. The flat plates (United McGill Corporation patents) increase the average electric field that can be used to collect the particles, and they provide an increased surface area for the collection of particles. Corona cannot be generated on flat plates by themselves, so corona-generating electrodes are placed ahead of and sometimes behind the flat plate collecting zones. These electrodes may be sharp-pointed needles attached to the edges of the plates or independent corona wires. Unlike plate-wire or tubular ESPs, this design operates equally well with either negative or positive polarity. The manufacturer has chosen to use positive polarity to reduce ozone generation.

A flat plate ESP operates with little or no corona current flowing through the collected dust, except directly under the corona needles or wires. This has two consequences. The first is that the unit is somewhat less susceptible to back corona than conventional units are because no back corona is generated in the collected dust, and particles charged with both polarities of ions have large collection surfaces available. The second consequence is that the lack of current in the collected layer causes an electrical force that tends to remove the layer from the collecting surface; this can lead to high rapping losses.

Flat plate ESPs seem to have wide application for high-resistivity particles with small (1 to 2 μm) mass median diameters (MMDs). These applications especially emphasize the strengths of the design because the electrical dislodging forces are weaker for small particles than for large ones. Fly ash has been successfully collected with this type of ESP, but low-flow velocity appears to be critical for avoiding high rapping losses.

6.1.2.3 Tubular Precipitators

The original ESPs were tubular like the smokestacks they were placed on, with the high-voltage electrode running along the axis of the tube. Tubular precipitators have typical applications in sulfuric acid plants, coke oven by-product gas cleaning (tar removal), and, recently, iron and steel sinter plants. Such tubular units are still used for some applications, with many tubes operating in parallel to handle increased gas flows. The tubes may be formed as a circular, square, or hexagonal honeycomb with gas flowing upwards or downwards. The length of the tubes can be selected to fit conditions. A tubular ESP can be tightly sealed to prevent leaks of material, especially valuable or hazardous material.

A tubular ESP is essentially a one-stage unit and is unique in having all the gas pass through the electrode region. The high-voltage electrode operates at one voltage for the entire length of the tube, and the current varies along the length as the particles are removed from the system. No sneakage paths are around the collecting region, but corona nonuniformities may allow some particles to avoid charging for a considerable fraction of the tube length.

Tubular ESPs comprise only a small portion of the ESP population and are most commonly applied where the particulate is either wet or sticky. These ESPs, usually cleaned with water, have reentrainment losses of a lower magnitude than do the dry particulate precipitators.

6.1.2.4 Wet Precipitators

Any of the precipitator configurations discussed above may be operated with wet walls instead of dry. The water flow may be applied intermittently or continuously to wash the collected particles into a sump for disposal. The advantage of the wet wall precipitator is that it has no problems with rapping reentrainment or with back corona. The disadvantage is the increased complexity of the wash and the fact that the collected slurry must be handled more carefully than a dry product, adding to the expense of disposal.

6.1.2.5 Two-Stage Precipitators

The previously described precipitators are all parallel in nature; *i.e.*, the discharge and collecting electrodes are side by side. The two-stage precipitator invented by Penney is a series device with the discharge electrode, or ionizer, preceding the collector electrodes. For indoor applications, the unit is operated with positive polarity to limit ozone generation.

Advantages of this configuration include more time for particle charging,

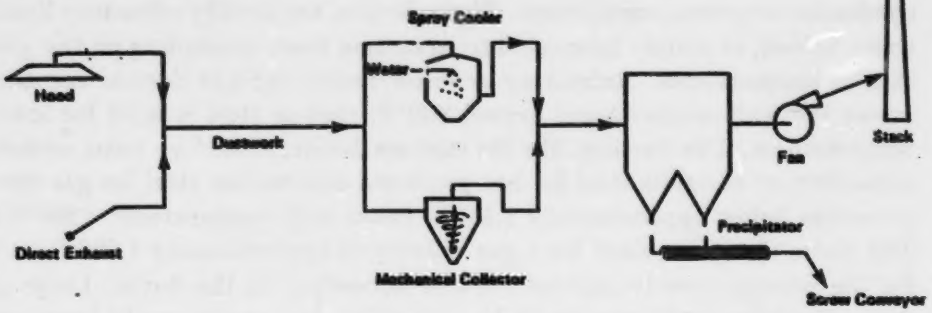


Figure 6.3: Control Device and Typical Auxilary Equipment

less propensity for back corona, and economical construction for small sizes. This type of precipitator is generally used for gas flow volumes of 50,000 acfm and less and is applied to submicrometer sources emitting oil mists, smokes, fumes, or other sticky particulates because there is little electrical force to hold the collected particulates on the plates. Modules consisting of a mechanical prefilter, ionizer, collecting-plate cell, after-filter, and power pack may be placed in parallel or series-parallel arrangements. Preconditioning of gases is normally part of the system. Cleaning may be by water wash of modules removed from the system up to automatic, in-place detergent spraying of the collector followed by air-blow drying.

Two-stage precipitators are considered to be separate and distinct types of devices compared to large, high-gas-volume, single-stage ESPs. The smaller devices are usually sold as pre-engineered, package systems.

6.1.3 Auxiliary Equipment

Typical auxiliary equipment associated with an ESP system is shown schematically in Figure 6.3. Along with the ESP itself, a control system usually includes the following auxiliary equipment: a capture device (*i. e.*, hood or direct exhaust connection); ductwork; dust removal equipment (screw conveyor, etc.); fans, motors, and starters; and a stack. In addition, spray coolers and mechanical collectors may be needed to precondition the gas before it reaches the ESP. Capture devices are usually hoods that exhaust pollutants into the ductwork or are direct exhaust couplings attached to a combustor or process equipment. These devices are usually refractory lined, water cooled, or simply fabricated from carbon steel, depending on the gas-stream temperatures. Refractory or water-cooled capture devices are used where the wall temperatures exceed 800°F; carbon steel is used for lower temperatures. The ducting, like the capture device, should be water cooled, refractory, or stainless steel for hot processes and carbon steel for gas temperatures below approximately 1,150°F (duct wall temperatures <800°F). The ducts should be sized for a gas velocity of approximately 4,000 ft/min for the average case to prevent particle deposition in the ducts. Large or dense particles might require higher velocities, but rarely would lower velocities be used. Spray chambers may be required for processes where the addition of moisture, or decreased temperature or gas volume, will improve precipitation or protect the ESP from warpage. For combustion processes with exhaust gas temperatures below approximately 700°F, cooling would not be required, and the exhaust gases can be delivered directly to the precipitator.

When much of the pollutant loading consists of relatively large particles, mechanical collectors, such as cyclones, may be used to reduce the load on the ESP, especially at high inlet concentrations. The fans provide the motive power for air movement and can be mounted before or after the ESP. A stack, normally used, vents the cleaned stream to the atmosphere. Screw conveyors or pneumatic systems are often used to remove captured dust from the bottom of the hoppers.

Wet ESPs require a source of wash water to be injected or sprayed near the top of the collector plates either continuously or at timed intervals. The water flows with the collected particles into a sump from which the fluid is pumped. A portion of the fluid may be recycled to reduce the total amount of water required. The remainder is pumped directly to a settling pond or passed through a dewatering stage, with subsequent disposal of the sludge.

Gas conditioning equipment to improve ESP performance by changing dust resistivity is occasionally used as part of the original design, but more frequently it is used to upgrade existing ESPs. The equipment injects an agent into the gas stream ahead of the ESP. Usually, the agent mixes with the particles and alters their resistivity to promote higher migration velocity, and thus higher collection efficiency. However, electrical properties of the gas may change, rather than dust resistivity. For instance, cooling the gas will allow more voltage to be applied before sparking occurs. Significant conditioning agents that are used include SO_3, H_2SO_4, sodium compounds, ammonia, and water, but the major conditioning agent by usage is SO_3. A typical dose rate for any of the gaseous agents is 10 to 30 ppm by volume.

The equipment required for conditioning depends on the agent being used. A typical SO_3 conditioner requires a supply of molten sulfur. It is stored in a heated vessel and supplied to a burner, where it is oxidized to SO_2. The SO_2 gas is passed over a catalyst for further oxidation to SO_3. The SO_3 gas is then injected into the flue gas stream through a multi-outlet set of probes that breach a duct. In place of a sulfur burner to provide SO_2, liquid SO_2 may be vaporized from a storage tank. Although their total annual costs are higher, the liquid SO_2 systems have lower capital costs and are easier to operate than the molten sulfur based systems.

Water or ammonia injection requires a set of spray nozzles in the duct, along with pumping and control equipment.

Sodium conditioning is often done by coating the coal on a conveyor with a powder compound or a water solution of the desired compound. A hopper or storage tank is often positioned over the conveyor for this purpose.

6.1.4 Electrostatic Precipitation Theory

The theory of ESP operation requires many scientific disciplines to describe it thoroughly. The ESP is basically an electrical machine. The principal actions are the charging of particles and forcing them to the collector plates. The amount of charged particulate matter affects the electrical operating point of the ESP. The transport of the particles is affected by the level of turbulence in the gas. The losses mentioned earlier, sneakage and rapping reentrainment, are major influences on the total performance of the system. The particle properties also have a major effect on the operation of the unit.

The following subsections will explain the theory behind (1) electrical operating points in the ESP, (2) particle charging, (3) particle collection, and (4) sneakage and rapping reentrainment. General references for these topics are White [1] or Lawless and Sparks [2].

6.1.4.1 Electrical Operating Point

The electrical operating point of an ESP section is the value of voltage and current at which the section operates. As will become apparent, the best collection occurs when the highest electric field is present, which roughly corresponds to the highest voltage on the electrodes. In this work, the term "section" represents one set of plates and electrodes in the direction of flow. This unit is commonly called a "field", and a "section" or "bus section" represents a subdivision of a "field" perpendicular to the direction of flow. In an ESP model and in sizing applications, the two terms "section" and "field" are used equivalently because the subdivision into bus sections should have no effect on the model. This terminology has probably arisen because of the frequent use of the word "field" to refer to the electric field.

The lowest acceptable voltage is the voltage required for the formation of a corona, the electrical discharge that produces ions for charging particles. The (negative) corona is produced when an occasional free electron near the high-voltage electrode, produced by a cosmic ray, gains enough energy from the electric field to ionize the gas and produce more free electrons. The electric field for which this process is self-sustained has been determined experimentally. For round wires, the field at the surface of the wire is given by:

$$E_c = 3.126 \times 10^6 d_r [1 + 0.0301 (d_r / r_w)^{0.5}] \tag{6.1}$$

where

E_c = corona onset field *at the wire surface* (V/m)
d_r = relative gas density, referred to 1 atm pressure and 20°C (dimensionless)
r_w = radius of the wire, meters (m)

This is the field required to produce "glow" corona, the form usually seen in the laboratory on smooth, clean wires. The glow appears as a uniform, rapidly moving diffuse light around the electrode. After a period of operation, the movement concentrates into small spots on the wire surface, and the corona assumes a tuft-like appearance. The field required to produce "tuft" corona, the form found in full-scale ESPs, is 0.6 times the value of E_c.

The voltage that must be applied to the wire to obtain this value of field, V_c, is found by integrating the electric field from the wire to the plate. The field follows a simple "$1/r$" dependence in cylindrical geometry. This leads to a logarithmic dependence of voltage on electrode dimensions. In the plate-wire geometry, the field dependence is somewhat more complex, but the voltage still shows the logarithmic dependence. V_c is given by:

$$V_c = E_c r_w \ln\left(\frac{d}{r_w}\right) \tag{6.2}$$

where

V_c = corona onset voltage (V)

$$d = \begin{cases} \text{outer cylinder radius for tubular ESP (m)} \\ 4/\pi \times \text{(wire-plate separation) for plate-wire ESP (m)} \end{cases}$$

No current will flow until the voltage reaches this value, but the amount of current will increase steeply for voltages above this value. The maximum current density (amperes/square meter) on the plate or cylinder directly under the wire is given by:

$$j = \mu\epsilon\frac{V^2}{L^3} \tag{6.3}$$

where

j = maximum current density (A/m^2)
μ = ion mobility (m^2/Vs) (meter^2/volt second)
ϵ = free space permittivity (8.845×10^{-12} F/m)(Farad/meter)
V = applied voltage (V)
L = shortest distance from wire to collecting surface (m)

For tuft corona, the current density is zero until the corona onset voltage is reached, when it jumps almost to this value of j within a few hundred volts, directly under a tuft.

The region near the wire is strongly influenced by the presence of ions there, and the corona onset voltage magnitude shows strong spatial variations. Outside the corona region, it is quite uniform.

The electric field is strongest along the line from wire to plate and is approximated very well, except near the wire, by:

$$E_{max} = V/L \tag{6.4}$$

where

E_{max} = maximum field strength (V/m)

When the electric field throughout the gap between the wire and the plate becomes strong enough, a spark will occur, and the voltage cannot be increased without severe sparking occurring. The field at which sparking occurs is not sharply defined, but a reasonable value is given by:

$$E_s = 6.3 \times 10^5 \left(\frac{273}{T}P\right)^{1.65} \tag{6.5}$$

where

E_s = sparking field strength (V/m)
T = absolute temperature (K)
P = gas pressure (atm)

This field would be reached at a voltage of, for example, 35,000 V for a plate-wire spacing of 11.4 cm (4.5 in.) at a temperature of 149°C (300°F). The ESP will generally operate near this voltage in the absence of back corona. E_{max} will be equal to or less than E_s.

Instead of sparking, back corona may occur if the electric field in the dust layer, resulting from the current flow in the layer, reaches a critical value of about 1×10^6 V/m. Depending on conditions, the back corona may enhance sparking or may generate so much current that the voltage cannot be raised any higher. The field in the layer is given by:

$$E_l = j\rho \tag{6.6}$$

where

E_l = electric field in dust layer (V/m)
ρ = resistivity of the collected material (ohm-m)

6.1.4.2 Particle Charging

Charging of particles takes place when ions bombard the surface of a particle. Once an ion is close to the particle, it is tightly bound because of the image charge within the particle. The "image charge" is a representation of the charge distortion that occurs when a real charge approaches a conducting surface. The distortion is equivalent to a charge of opposite magnitude to the real charge, located as far below the surface as the real charge is above it. The notion of the fictitious charge is similar to the notion of an image in a mirror, hence the name. As more ions accumulate on a particle, the total charge tends to prevent further ionic bombardment.

There are two principal charging mechanisms: diffusion charging and field charging. Diffusion charging results from the thermal kinetic energy of the ions overcoming the repulsion of the ions already on the particle. Field charging occurs when ions follow electric field lines until they terminate on a particle. In general, both mechanisms are operative for all sizes of particles. Field charging, however, adds a larger percentage of charge on particles greater than about 2 μm in diameter, and diffusion charging adds a greater percentage on particles smaller than about 0.5 μm.

Diffusion charging, as derived by White [1], produces a logarithmically increasing level of charge on particles, given by:

$$q(t) = \left(\frac{rkT}{e}\right) \ln(1 + \tau) \tag{6.7}$$

where

$q(t)$ = particle charge (C) as function of time, t, in seconds
r = particle radius (m)
k = Boltzmann's constant (j/K)
T = absolute temperature (K)
e = electron charge (1.67×10^{-19} C)
τ = dimensionless time given by:

$$\tau = \frac{\pi r v N e^2 \theta}{kT} \tag{6.8}$$

where

v = mean thermal speed of the ions (m/s)
N = ion number concentration near the particle (No./m^3)
θ = real time (exposure time in the charging zone) (s)

Diffusion charging never reaches a limit, but it becomes very slow after about three dimensionless time units. For fixed exposure times, the charge on a particle is proportional to its radius.

Field charging also exhibits a characteristic time-dependence, given by:

$$q(t) = q_s\theta/(\theta + \tau') \tag{6.9}$$

where

q_s = saturation charge, charge at infinite time (C)
θ = real time (s)
τ' = another dimensionless time unit

The saturation charge is given by:

$$q_s = 12\pi\epsilon r^2 E \tag{6.10}$$

where

ϵ = free space permittivity (F/m)
E = external electric field applied to the particle (V/m)

The saturation charge is proportional to the square of the radius, which explains why field charging is the dominant mechanism for larger particles. The field charging time constant is given by:

$$\tau' = 4\epsilon/Ne\mu \tag{6.11}$$

where

μ = ion mobility

Strictly speaking, both diffusion and field charging mechanisms operate at the same time on all particles, and neither mechanism is sufficient to explain the charges measured on the particles. It has been found empirically that a very good approximation to the measured charge is given by the sum of the charges predicted by equations 6.7 and 6.9 independently of one another:

$$q_{tot}(t) = q_d(t) + q_f(t) \tag{6.12}$$

where

$q_{tot}(t)$ = particle charge due to both mechanisms
$q_d(t)$ = particle charge due to diffusion charging
$q_f(t)$ = particle charge due to field charging

6.1.4.3 Particle Collection

The electric field in the collecting zone produces a force on a particle proportional to the magnitude of the field and to the charge:

$$F_e = qE \tag{6.13}$$

where

$\mathbf{F}_e$ = force due to electric field (N)
q = charge on particle (C)
E = electric field (V/m)

Because the field charging mechanism gives an ultimate charge proportional to the electric field, the force on large particles is proportional to the square of the field, which shows the advantage for maintaining as high a field as possible.

The motion of the particles under the influence of the electric field is opposed by the viscous drag of the gas. By equating the electric force and the drag force component due to the electric field (according to Stokes' law), we can obtain the particle velocity:

$$v(q,E,r) = \frac{q(E,r) \times E \times C(r)}{6\pi\eta r} \tag{6.14}$$

where

$v(q,E,r)$ = particle velocity (m/s)
$q(E,r)$ = particle charge (C)
$C(r)$ = Cunningham correction to Stokes' law (dimensionless)
η = gas viscosity (kg/ms)

The particle velocity is the rate at which the particle moves along the electric field lines, *i.e.*, toward the walls.

For a given electric field, this velocity is usually at a minimum for particles of about 0.5 μm diameter. Smaller particles move faster because the charge does not decrease very much, but the Cunningham factor increases rapidly as radius decreases. Larger particles have a charge increasing as r^2 and a viscous drag only increasing as r^1; the velocity then increases as r.

Equation 6.14 gives the particle velocity with respect to still air. In the ESP, the flow is usually very turbulent, with instantaneous gas velocities of the same magnitude as the particles velocities, but in random directions. The motion of particles toward the collecting plates is therefore a statistical process, with an average component imparted by the electric field and a fluctuating component from the gas turbulence.

This statistical motion leads to an exponential collection equation, given by:

$$N(r) = N_0(r) \times \exp(-v(r)/v_0) \tag{6.15}$$

where

$N(r)$ = particle concentration of size r at the exit of the collecting zone (No./m^3)

$N_0(r)$ = particle concentration of size r at the entrance of the zone (No./m^3)

$v(r)$ = size-dependent particle velocity (m/s)

v_0 = characteristic velocity of the ESP (m/s), given by:

$$v_0 = Q/A = 1/\mathrm{SCA} \tag{6.16}$$

where

Q = volume flow rate of the gas (m^3/s)

A = plate area for the ESP collecting zone (m^2)

SCA = specific collection area (A/Q) (s/m)

When this collection equation is averaged over all the particle sizes and weighted according to the concentration of each size, the Deutsch equation results, with the penetration (fraction of particles escaping) given by:

$$p = \exp(-w_e \times \mathrm{SCA}) \tag{6.17}$$

where

p = penetration (fraction)

w_e = effective migration velocity for the particle ensemble (m/s)

The efficiency is given by:

$$\mathrm{Eff}(\%) = 100(1 - p) \tag{6.18}$$

and is the number most often used to describe the performance of an ESP.

6.1.4.4 Sneakage and Rapping Reentrainment

Sneakage and rapping reentrainment are best considered on the basis of the sections within an ESP. Sneakage occurs when a part of the gas flow bypasses the collection zone of a section. Generally, the portion of gas that bypasses the zone is thoroughly mixed with the gas that passes through the zone before all the gas enters the next section. This mixing cannot always be assumed, and when sneakage paths exist around several sections, the performance of the whole ESP is seriously affected. To describe the effects of sneakage and rapping reentrainment mathematically, we first consider sneakage by itself and then consider the effects of rapping as an average over many rapping cycles.

On the assumption that the gas is well mixed between sections, the penetration for each section can be expressed as:

$$p_s = S_N + [(1 - S_N) \times p_c(Q')] \tag{6.19}$$

where

p_s = section's fractional penetration
S_N = fraction of gas bypassing the section (sneakage)
$p_c(Q')$ = fraction of particles penetrating the collection zone, which is functionally dependent on Q′, the gas volume flow in the collection zone, reduced by the sneakage (m^3/s)

The penetration of the entire ESP is the product of the section penetrations. The sneakage sets a lower limit on the penetration of particles through the section.

To calculate the effects of rapping, we first calculate the amount of material captured on the plates of the section. The fraction of material that was caught is given by:

$$m/m_0 = 1 - p_s = 1 - S_N - [(1 - S_N) \times p_c(Q')] \tag{6.20}$$

where

m/m_0 = mass fraction collected from the gas stream

This material accumulates until the plates are rapped, whereupon most of the material falls into the hopper for disposal, but a fraction of it is reentrained and leaves the section. Experimental measurements have been conducted on fly ash ESPs to evaluate the fraction reentrained, which averages about 12 percent.

The average penetration for a section, including sneakage and rapping reentrainments, is:

$$p_s = S_N + [(1 - S_N) \times p_c(Q')] + RR(1 - S_N)[1 - p_c(Q')] \tag{6.21}$$

where

RR = fraction reentrained

This can be written in a more compact form as:

$$p_e = LF + [(1 - LF) \times p_c(Q')] \tag{6.22}$$

by substituting LF (loss factor) for $S_N + RR(1 - S_N)$. These formulas can allow for variable amounts of sneakage and rapping reentrainment for each section, but there is no experimental evidence to suggest that it is necessary.

Fly ash precipitators analyzed in this way have an average S_N of 0.07 and an RR of 0.12. These values are the best available at this time, but some wet ESPs, which presumably have no rapping losses, have shown S_N values of 0.05 or less. These values offer a means for estimating the performance of ESPs whose actual characteristics are not known, but about which general statements can be made. For instance, wet ESPs would be expected to have RR = 0, as would ESPs collecting wet or sticky particles. Particulate materials with a much smaller mass mean diameter, MMD, than fly ash would be expected to have a lower RR factor because they are held more tightly to the plates and each other. Sneakage factors are harder to account for; unless special efforts have been made in the design to control sneakage, the 0.07 value should be used.

6.2 ESP Design Procedure

6.2.1 Specific Collecting Area

Specific collecting area (SCA) is a parameter used to compare ESPs and roughly estimate their collection efficiency. SCA is the total collector plate area divided by gas volume flow rate and has the units of s/m or s/ft. Since SCA is the ratio of A/Q, it is often expressed as $m^2/(m^3/s)$ or $ft^2/kacfm$, where kacfm is thousand acfm. SCA is also one of the most important factors in determining the capital and several of the annual costs (for example, maintenance and dust disposal costs) of the ESP because it determines the size of the unit. Because of the various ways in which SCA can be expressed, Table 6.1 gives equivalent SCAs in the different units for what would be considered a small, medium, and large SCA.

Table 6.1: Small, Medium, and Large SCAs as Expressed by Various Units

Units	Small	Medium	Large
ft^2/kacfm	100	400	900
s/m	19.7	78.8	177
s/ft	6	24	54

5.080 ft^2/kacfm = 1 (s/m).

The design procedure is based on the loss factor approach of Lawless and Sparks [2] and considers a number of process parameters. It can be calculated by hand, but it is most conveniently used with a spreadsheet program. For many uses, tables of effective migration velocities can be used to obtain the SCA required for a given efficiency. In the following subsection, tables have been calculated using the design procedure for a number of different particle sources and for differing levels of efficiency. If a situation is encountered that is not covered in these tables, then the full procedure that appears in the subsequent subsection should be used.

6.2.1.1 SCA Procedure with Known Migration Velocity

If the migration velocity is known, then equation 6.17 can be rearranged to give the SCA:

$$\text{SCA} = -\ln(p)/w_e \tag{6.23}$$

A graphical solution to equation 6.23 is given in Figure 6.4. The migration velocities have been calculated for three main precipitator types: plate-wire, flat plate, and wet wall ESPs of the plate-wire type. The following three tables, keyed to design efficiency as an easily quantified variable, summarize the migration velocities under various conditions:

- In Table 6.2, the migration velocities are given for a plate-wire ESP with conditions of no back corona and severe back corona; temperatures appropriate for each process have been assumed.

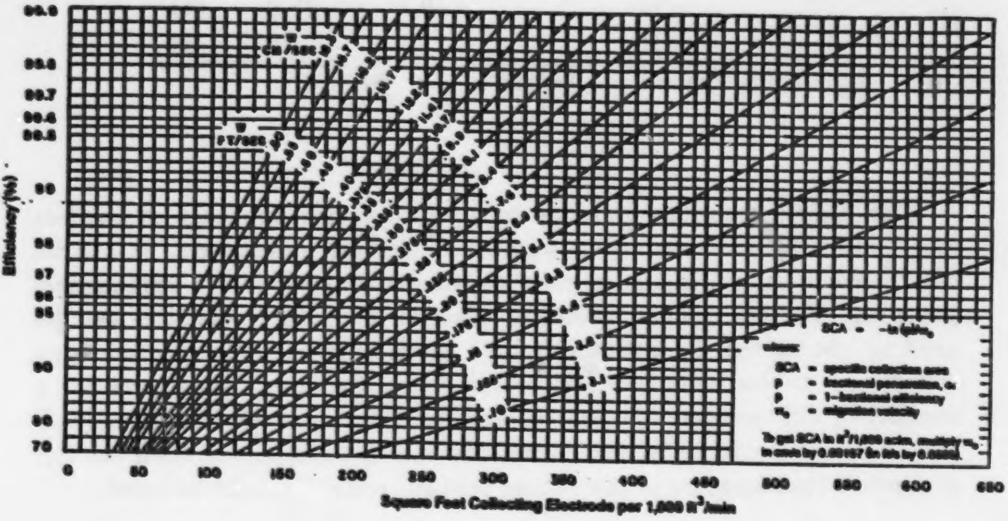


Figure 6.4: Chart for Finding SCA

- **In Table 6.3, the migration velocities calculated for a wet wall ESP of the plate-wire type assume no back corona and no rapping reentrainment.**
- **In Table 6.4, the flat plate ESP migration velocities are given only for no back corona conditions because they appear to be less affected by high-resistivity dusts than the plate-wire types.**

It is generally expected from experience that the migration velocity will decrease with increasing efficiency. In Tables 6.2 through 6.4, however, the migration velocities show some fluctuations. This is because the number of sections must be increased as the efficiency increases, and the changing sectionalization affects the overall migration velocity. This effect is particularly noticeable, for example, in Table 6.4 for glass plants. When the migration velocities in the tables are used to obtain SCAs for the different efficiencies in the tables, the SCAs will increase as the efficiency increases.

6.2.1.2 Full SCA Procedure

The full procedure for determining the SCA for large plate-wire, flat plate, and (with restrictions) tubular dry ESPs is given here. This procedure does not apply to the smaller, two-stage precipitators because these are packaged modules generally sized and sold on the basis of the waste gas volumetric flow rate. Nor does this procedure apply to determining the SCA for wet ESPs. The full procedure consists of the 15 steps given below:

Step 1 - Determine the design efficiency, Eff(%). Efficiency is the most commonly used term in the industry and is the reference value for guarantees; however, if it has not been specified, it can be computed as follows:

$$\text{Eff}(\%) = 100 \times (1 - \text{outlet load}/\text{inlet load})$$

Step 2 - Compute design penetration, p:

$$p = 1 - (\text{Eff}/100)$$

Table 6.2: Plate-wire ESP Migration Velocities (cm/s)[a]

		Design Efficiency, %			
Particle Source		95	99	99.5	99.9
Bituminous coal fly ash[b]	(no BC)	12.6	10.1	9.3	8.2
	(BC)	3.1	2.5	2.4	2.1
Sub-bituminous coal fly ash in tangential-fired boiler[b]	(no BC)	17.0	11.8	10.3	8.8
	(BC)	4.9	3.1	2.6	2.2
Other coal[b]	(no BC)	9.7	7.9	7.9	7.2
	(BC)	2.9	2.2	2.1	1.9
Cement kiln[c]	(no BC)	1.5	1.5	1.8	1.8
	(BC)	0.6	0.6	0.5	0.5
Glass plant[d]	(no BC)	1.6	1.6	1.5	1.5
	(BC)	0.5	0.5	0.5	0.5
Iron/steel sinter plant dust with mechanical precollector[b]	(no BC)	6.8	6.2	6.6	6.3
	(BC)	2.2	1.8	1.8	1.7
Kraft-paper recovery boiler[b]	(no BC)	2.6	2.5	3.1	2.9
Incinerator fly ash[e]	(no BC)	15.3	11.4	10.6	9.4
Copper reverberatory furnace[f]	(no BC)	6.2	4.2	3.7	2.9
Copper converter[g]	(no BC)	5.5	4.4	4.1	3.6
Copper roaster[h]	(no BC)	6.2	5.5	5.3	4.8
Coke plant combustion stack[i]	(no BC)	1.2[j]	—	—	—

BC = Back corona.

[a] To convert cm/s to ft/s, multiply cm/s by 0.0328. Computational procedure uses SI units, to convert cm/s to m/s, multiply cm/s by 0.01. Assumes same particle size as given in full computational procedure.

[b] At 300°F. Depending on individual furnace/boiler conditions, chemical nature of the fly ash, and availability of naturally occurring conditioning agents (*e.g.*, moisture in the gas stream), migration velocities may vary considerably from these values. Likely values are in the range from back corona to no back corona.

[c] At 600°F. [d] At 500°F. [e] At 250°F. [f] 450 to 570 °F.

[g] 500 to 700 °F. [h] 600 to 660 °F. [i] 360 to 450 °F.

[j] Data available only for inlet concentrations in the range of 0.02 to 0.2 g/s m^3 and for efficiencies less than 90 percent.

Table 6.3: Wet Wall Plate-wire ESP Migration Velocities

(No back corona, cm/s)[a]

Particle Source[b]	Design Efficiency, % 95	99	99.5	99.9
Bituminous coal fly ash	31.4	33.0	33.8	24.9
Sub-bituminous coal fly ash in tangential-fired boiler	40.0	42.7	44.1	31.4
Other coal	21.1	21.4	21.5	17.0
Cement kiln	6.4	5.6	5.0	5.7
Glass plant	4.6	4.5	4.3	3.8
Iron/steel sinter plant dust with mechanical precollector	14.0	13.7	13.3	11.6

[a] To convert cm/s to ft/s, multiply cm/s by 0.0328. Computational procedure uses SI units; to convert cm/s to m/s, multiply cm/s by 0.01. Assumes same particle size as given in full computational procedure.

[b] All sources assumed at 200 °F.

Table 6.4: Flat Plate ESP Migration Velocities[a]

(No back corona, cm/s)[b]

	Design Efficiency, %			
Particle Source	95	99	99.5	99.9
Bituminous coal fly ash[c]	13.2	15.1	18.6	16.0
Sub-bituminous coal fly ash in tangential-fired boiler[c]	28.6	18.2	21.2	17.7
Other coal[c]	15.5	11.2	15.1	13.5
Cement kiln[d]	2.4	2.3	3.2	3.1
Glass plant[e]	1.8	1.9	2.6	2.6
Iron/steel sinter plant dust with mechanical precollector[c]	13.4	12.1	13.1	12.4
Kraft-paper recovery boiler[c]	5.0	4.7	6.1	5.8
Incinerator fly ash[f]	25.2	16.9	21.1	18.3

[a] Assumes same particle size as given in full computational procedure. These values give the grounded collector plate SCA, from which the collector plate area is derived. In flat plate ESPs, the discharge or high-voltage plate area is typically 40 percent of the ground-plate area. The flat-plate manufacturer usually counts all the plate area (collector plates plus discharge plates) in meeting an SCA specification, which means that the velocities tabulated above must be divided by 1.4 to be used on the manufacturer's basis.

[b] To convert cm/s to ft/s, multiply cm/s by 0.0328. Computational procedure uses SI units; to convert cm/s to m/s, multiply cm/s by 0.01.

[c] At 300°F.

[d] At 600°F.

[e] At 500°F.

[f] At 250°F.

Step 3 - Compute or obtain the operating temperature, Tk, K. Temperature in Kelvin is required in the calculations which follow.

Step 4 - Determine whether severe back corona is present. Severe back corona usually occurs for dust resistivities above 2×10^{11} ohm-cm. Its presence will greatly increase the size of the ESP required to achieve a certain efficiency.

Step 5 - Determine the MMD of the inlet particle distribution MMD_i (μm). If this is not known, assume a value from the following table:

Source	MMD_i (μm)
Bituminous coal	16
Sub-bituminous coal, tangential boiler	21
Sub-bituminous coal, other boiler types	10 to 15
Cement kiln	2 to 5
Glass plant	1
Wood burning boiler	5
Sinter plant,	50
with mechanical precollector	6
Kraft Process Recovery	2
Incinerators	15 to 30
Copper reverberatory furnace	1
Copper converter	1
Coke plant combustion stack	1
Unknown	1

Step 6 - Assume values for sneakage, S_N, and rapping reentrainment, RR, from the following table:

ESP Type	S_N
Plate-wire	0.07
Wet wall	0.05
Flat plate	0.10

ESP/Ash Type	RR
Coal fly ash, or not known	0.124
Wet wall	0.0
Flat plate with gas velocity >1.5 m/s (not glass or cement)	0.15
Glass or cement	0.10

Step 7 - Assume values for the most penetrating size, MMD_p, and rapping puff size, MMD_r:

$$MMD_p = 2\ \mu m$$
$$MMD_r = 5\ \mu m \quad \text{for ash with } MMD_i > 5\ \mu m$$
$$MMD_r = 3\ \mu m \quad \text{for ash with } MMD_i < 5\ \mu m$$

where

MMD_p = the MMD of the size distribution emerging from a very efficient collecting zone

MMD_r = the MMD of the size distribution of rapped/reentrained material.

Step 8 - Use or compute the following factors for pure air:

$\epsilon 0 = 8.845\times10^{-12}$ free space permittivity (F/m)

$\eta = 1.72\times10^{-5}(Tk/273)^{0.71}$ gas viscosity (kg/ms)

$E_{bd} = 630{,}000\ (273/Tk)^{1.65}$ electric field at sparking (V/m)

$LF = S_N + RR(1 - S_N)$ loss factor (dimensionless)

For plate-wire ESPs:

$E_{avg} = E_{bd}/1.75$ average field with no back corona

$E_{avg} = 0.7 \times E_{bd}/1.75$ average field with severe back corona

For flat plate ESPs:

$E_{avg} = E_{bd} \times 5/6.3$ average field, no back corona, positive polarity

$E_{avg} = 0.7 \times E_{bd} \times 5/6.3$ average field, severe back corona, positive polarity

Step 9 - Assume the smallest number of sections for the ESP, n, such that $LF^n < p$. Suggested values of n are:

Eff(%)	n
<96.5	2
<99	3
<99.8	4
<99.9	5
>99.9	6

These values are for an LF of 0.185, corresponding to a coal fly ash precipitator. The values are approximate, but the best results are for the smallest allowable n.

Step 10 - Compute the average section penetration, p_s:

$$p_s = p^{1/n}$$

Step 11 - Compute the section collection penetration, p_c:

$$p_c = \frac{p_s - LF}{1 - LF}$$

If the value of n is too small, then this value will be negative and n must be increased.

Step 12 - Compute the particle size change factors, D and MMD_{rp}, which are constants used for computing the change of particle size from section to section:

$$D = p_s = S_N + p_c(1 - S_N) + RR(1 - S_N)(1 - p_c)$$

$$MMD_{rp} = RR(1 - S_N)(1 - p_c)MMD_r/D$$

Step 13 - Compute a table of particle sizes for sections 1 through n:

Section	MMD
1	$\mathrm{MMD}_1 = \mathrm{MMD}_i$
2	$\mathrm{MMD}_2 = \{\mathrm{MMD}_1 \times \mathrm{S}_N + [(1 - p_c) \times \mathrm{MMD}_p + p_c \times \mathrm{MMD}_1] \times p_c\}/\mathrm{D} + \mathrm{MMD}_{rp}$
3	$\mathrm{MMD}_3 = \{\mathrm{MMD}_2 \times \mathrm{S}_N + [(1 - p_c) \times \mathrm{MMD}_p + p_c \times \mathrm{MMD}_2] \times p_c\}/\mathrm{D} + \mathrm{MMD}_{rp}$
$\vdots$	$\vdots$
n	$\mathrm{MMD}_n = \{\mathrm{MMD}_{n-1} \times \mathrm{S}_N + [(1 - p_c) \times \mathrm{MMD}_p + p_c \times \mathrm{MMD}_{n-1}] \times p_c\}/\mathrm{D} + \mathrm{MMD}_{rp}$

Step 14 - Calculate the SCA for sections 1 through n, using MMD_n, η, ϵ, E_{avg}, and p_c:

$$\begin{aligned} \mathrm{SCA}_1 &= -(\eta/\epsilon) \times (1 - \mathrm{S}_N) \times \ln(\mathrm{p}_c)/(\mathrm{E}^2_{avg} \times \mathrm{MMD}_1 \times 10^{-6}) \\ &\vdots \\ \mathrm{SCA}_n &= -(\eta/\epsilon) \times (1 - \mathrm{S}_N) \times \ln(\mathrm{p}_c)/(\mathrm{E}^2_{avg} \times \mathrm{MMD}_n \times 10^{-6}) \end{aligned}$$

where the factor 10^{-6} converts micrometers to meters. Note that the only quantity changing in these expressions is MMD_x; therefore, the following relation can be used:

$$\mathrm{SCA}_{n+1} = \mathrm{SCA}_n \times \mathrm{MMD}_n/\mathrm{MMD}_{n+1}$$

Step 15 - Calculate the total SCA and the English SCA, ESCA:

$$\mathrm{SCA}\ (\mathrm{s/m}) = \sum_{i=1}^{n} \mathrm{SCA}_i$$

$$\text{ESCA (ft}^2/\text{kacfm}) = 5.080 \times \text{ SCA (s/m)}$$

This sizing procedure works best for p_e values less than the value of LF, which means the smallest value of n. Any ESP model is sensitive to the values of particle diameter and electric field. This one shows the same sensitivity, but the expressions for electric field are based on theoretical and experimental values. The SCA should not be strongly affected by the number of sections chosen; if more sections are used, the SCA of each section is reduced.

6.2.1.3 Specific Collecting Area for Tubular Precipitators

The procedure given above is suitable for large plate-wire or flat plate ESPs, but must be used with restrictions for tubular ESPs. Values of $S_N = 0.015$ and RR = 0 are assumed, and only one section is used.

Table 6.5 gives migration velocities that can be used with equation 6.23 to calculate SCAs for several tubular ESP applications.

6.2.2 Flow Velocity

A precipitator collecting a dry particulate material runs a risk of nonrapping (continuous) reentrainment if the gas velocity becomes too large. This effect is independent of SCA and has been learned through experience. For fly ash applications, the maximum acceptable velocity is about 1.5 m/s (5 ft/s) for plate-wire ESPs and about 1 m/s (3 ft/s) for flat plate ESPs. For low resistivity applications, design velocities of 3 ft/s or less are common to avoid nonrapping reentrainment. The frontal area of the ESP (W × H), *i.e.*, the area normal to the direction of gas flow, must be chosen to keep gas velocity low and to accommodate electrical requirements (*e.g.*, wire-to-plate spacing) while also ensuring that total plate area requirements are met. This area can be configured in a variety of ways. The plates can be short in height, long in the direction of flow, with several in parallel (making the width narrow). Or, the plates can be tall in height, short in the direction of flow, with many in parallel making the width large). After selecting a configuration, the gas velocity can be obtained by dividing the

Table 6.5: **Tubular ESP Migration Velocities**[a]

(cm/s)[b]

		Design Efficiency, %	
Particle Source		90	95
Cement kiln	(no BC)	2.2–5.4	2.1–5.1
	(BC)	1.1–2.7	1.0–2.6
Glass plant	(no BC)	1.4	1.3
	(BC)	0.7	0.7
Kraft-paper recovery boiler	(no BC)	4.7	4.4
Incinerator 15 μm MMD	(no BC)	40.8	39.
Wet, at 200°F MMD (μm)			
1		3.2	3.1
2		6.4	6.2
5		16.1	15.4
10		32.2	30.8
20		64.5	61.6

BC = Back corona

[a] These rates were calculated on the basis of:
$S_N = 0.015$, RR = 0, one section only.
These are in agreement with operating tubular ESPs; extension of results to more than one section is not recommended.

[b] To convert cm/s to ft/s, multiply cm/s by 0.0328.

volume flow rate, Q, by the frontal area of the ESP:

$$v_{gas} = \frac{Q}{WH} \tag{6.24}$$

where

v_{gas} = gas velocity (m/s)
W = width of ESP entrance (m)
H = height of ESP entrance (m)

When meeting the above restrictions, this value of velocity also ensures that turbulence is not strongly developed, thereby assisting in the capture of particles.

6.2.3 Pressure Drop Calculations

The pressure drop in an ESP is due to four main factors:

- Diffuser plate (usually present)—(perforated plate at the inlet)
- Transitions at the ESP inlet and outlet
- Collection plate baffles (stiffeners) or corrugations
- Drag of the flat collection plate

The total pressure drop is the sum of the individual pressure drops, but any one of these sources may dominate all other contributions to the pressure drop. Usually, the pressure drop is not a design-driving factor, but it needs to be maintained at an acceptably low value. Table 6.6 gives typical pressure drops for the four factors. The ESP pressure drop, usually less than about 0.5 in. H_2O, is much lower than for the associated collection system and ductwork. With the conveying velocities used for dust collected in ESPs, generally 4,000 ft/min or greater, system pressure drops are usually in the range of 2 to 10 in. H_2O, depending upon the ductwork length and configuration as well as the type(s) of preconditioning devices(s) used upstream.

The four main factors contributing to pressure drop are described briefly below.

Table 6.6: Components of ESP Pressure Drop

	Typical Pressure Drop (in. H_2O)	
Component	Low	High
Diffuser	0.010	0.09
Inlet transition	0.07	0.14
Outlet transition	0.007	0.015
Baffles	0.0006	0.123
Collection plates	0.0003	0.008
Total	0.09	0.38

The diffuser plate is used to equalize the gas flow across the face of the ESP. It typically consists of a flat plate covered with round holes of 5 to 7 cm diameter (2 to 2.5 in.) having an open area of 50 to 65 percent of the total. Pressure drop is strongly dependent on the percent open area, but is almost independent of hole size.

The pressure drop due to gradual enlargement at the inlet is caused by the combined effects of flow separation and wall friction and is dependent on the shape of the enlargement. At the ESP exit, the pressure drop caused by a short, well-streamlined gradual contraction is small.

Baffles are installed on collection plates to shield the collected dust from the gas flow and to provide a stiffening effect to keep the plates aligned parallel to one another. The pressure drop due to the baffles depends on the number of baffles, their protrusion into the gas stream with respect to electrode-to-plate distance, and the gas velocity in the ESP.

The pressure drop of the flat collection plates is due to friction of the gas dragging along the flat surfaces and is so small compared to other factors that it may usually be neglected in engineering problems.

6.2.4 Particle Characteristics

Several particle characteristics are important for particle collection. It is generally assumed that the particles are spherical or spherical enough to be described by some equivalent spherical diameter. Highly irregular or elongated particles may not behave in ways that can be easily described.

The first important characteristic is the mass of particles in the gas stream, *i.e.*, the particle loading. This quantity usually is determined by placing a filter in the gas stream, collecting a known volume of gas, and determining the weight gain of the filter. Because the ESP operates over a wide range of loadings as a constant efficiency device, the inlet loading will determine the outlet loading directly. If the loading becomes too high, the operation of the ESP will be altered, usually for the worse.

The second characteristic is the size distribution of the particles, often expressed as the cumulative mass less than a given particle size. The size distribution describes how many particles of a given size there are, which is important because ESP efficiency varies with particle size. In practical terms, an ESP will collect all particles larger than 10 μm in diameter better than ones smaller than 10 μm. Only if most of the mass in the particles is concentrated above 10 μm would the actual size distribution above 10 μm be needed.

In lieu of cumulative mass distributions, the size distribution is often described by log-normal parameters. That is, the size distribution appears as a probabilistic normal curve if the logarithm of particle size used is the abscissa. The two parameters needed to describe a log-normal distribution are the mass median (or mean) diameter and the geometric standard deviation.

The MMD is the diameter for which one-half of the particulate mass consists of smaller particles and one-half is larger (see the Procedure, Step 5, of Subsection 6.2.1.2). If the MMD of a distribution is larger than about 3 μm, the ESP will collect all particles larger than the MMD at least as well as a 3 μm particle, representing one-half the mass in the inlet size distribution.

The geometric standard deviation is the equivalent of the standard deviation of the normal distribution: It describes how broad the size distribution is. The geometric standard deviation is computed as the ratio of the diameter corresponding to 84 percent of the total cumulative mass to the MMD; it is always a number greater than 1. A distribution with particles of all the same size (monodisperse) has a geometric standard deviation of 1. Geometric standard deviations less than 2 represent rather narrow distributions. For combustion sources, the geometric standard deviations range from 3 to 5 and are commonly in the 3.5 to 4.5 range.

A geometric standard deviation of 4 to 5, coupled with an MMD of less than 5 μm, means that there is a substantial amount of submicrometer material. This situation may change the electrical conditions in an ESP by the phenomenon known as "space charge quenching", which results in high operating voltages but low currents. It is a sign of inadequate charging and reduces the theoretical efficiency of the ESP. This condition must be evaluated carefully to be sure of adequate design margins.

6.2.5 Gas Characteristics

The gas characteristics most needed for ESP design are the gas volume flow and the gas temperature. The volume flow, multiplied by the design SCA, gives the total plate area required for the ESP. If the volume flow is known at one temperature, it may be estimated at other temperatures by applying the ideal gas law. Temperature and volume uncertainties will outweigh inaccuracies of applying the ideal gas law.

The temperature of the gas directly affects the gas viscosity, which increases with temperature. Gas viscosity is affected to a lesser degree by the gas composition, particularly the water vapor content. In lieu of viscosity values for a particular gas composition, the viscosity for air may be used. Viscosity enters the calculation of SCA directly, as seen in Step 14 of the design procedure (page 6-33).

The gas temperature and composition can have a strong effect on the resistivity of the collected particulate material. Specifically, moisture and acid gas components may be chemisorbed on the particles in a sufficient amount to lower the intrinsic resistivity dramatically (orders of magnitude). For other types of materials, there is almost no effect. Although it is not possible to treat resistivity here, the designer should be aware of the potential sensitivity of the size of the ESP to resistivity and the factors influencing it.

The choice of power supplies' size (current capacity and voltage) to be used with a particular application may be influenced by the gas characteristics. Certain applications produce gas whose density may vary significantly from typical combustion sources (density variation may result from temperature, pressure, and composition). Gas density affects corona starting voltages and voltages at which sparking will occur.

6.2.6 Cleaning

Cleaning the collected materials from the plates often is accomplished intermittently or continuously by rapping the plates severely with automatic hammers or pistons, usually along their top edges, except in the case of wet ESPs that use water. Rapping dislodges the material, which then falls down the length of the plate until it lands in a dust hopper. The dust characteristics, rapping intensity, and rapping frequency determine how much of the material is reentrained and how much reaches the hopper permanently.

For wet ESPs, consideration must be given to handling waste waters. For simple systems with innocuous dusts, water with particles collected by the ESP may be discharged from the ESP system to a solids-removing clarifier (either dedicated to the ESP or part of the plant wastewater treatment system) and then to final disposal. More complex systems may require skimming and sludge removal, clarification in dedicated equipment, pH adjustment, and/or treatment to remove dissolved solids. Spray water from the ESP preconditioner may be treated separately from the water used to flood the ESP collecting plates, so that the cleaner of the two treated waters may be returned to the ESP. Recirculation of treated water to the ESP may approach 100 percent.

The hopper should be designed so that all the material in it slides to the very bottom, where it can be evacuated periodically, as the hopper becomes full. Dust is removed through a valve into a dust-handling system, such as a pneumatic conveyor. Hoppers often are supplied with auxiliary heat to prevent the formation of lumps or cakes and the subsequent blockage of the dust handling system.

6.2.7 Construction Features

The use of the term "plate-wire geometry" may be somewhat misleading. It could refer to three different types of discharge electrodes: weighted wires hung from a support structure at the top of the ESP, wire frames in which wires are strung tautly in a rigid support frame, or rigid electrodes constructed from a single piece of fabricated metal. In recent years, there has been a trend toward using wire frames or rigid discharge electrodes in place of weighted wire discharge electrodes (particularly in coal-fired boiler applications). This trend has been stimulated by the user's desire for increased ESP reliability. The wire frames and rigid electrodes are less prone to failure by breakage and are readily cleaned by impulse-type cleaning equipment.

Other differences in construction result from the choice of gas passage (flow lane) width or discharge electrode to collecting electrode spacing. Typically, discharge to collecting electrode spacing varies from 11 to 19 cm (4.3 to 7.5 in.). Having a large spacing between discharge and collecting electrodes allows higher electric fields to be used, which tends to improve dust collection. To generate larger electric fields, however, power supplies must produce higher operating voltages. Therefore, it is necessary to balance the cost savings achieved with larger electrode spacing against the higher cost of power supplies that produce higher operating voltages.

Most ESPs are constructed of mild steel. ESP shells are constructed typically of 3/16 or 1/4 in. mild steel plate. Collecting electrodes are generally fabricated from lighter gauge mild steel. A thickness of 18 gauge is common, but it will vary with size and severity of application.

Wire discharge electrodes come in varied shapes from round to square or barbed. A diameter of 2.5 mm (0.1 in.) is common for weighted wires, but other shapes used have much larger effective diameters, *e.g.*, 64 mm (0.25 in.) square electrodes.

Stainless steel may be used for corrosive applications, but it is uncommon except in wet ESPs. Stainless steel discharge electrodes have been found to be prone to fatigue failure in dry ESPs with impact-type electrode cleaning systems.[3]

Precipitators used to collect sulfuric acid mist in sulfuric acid plants are constructed of steel, but the surfaces in contact with the acid mist are lead-lined. Precipitators used on paper mill black liquor recovery boilers are steam-jacketed. Of these two, recovery boilers have by far the larger number of ESP applications.

Table 6.7: Standard Options for Basic Equipment

Option	Cost adder (%)
1 - Inlet and outlet nozzles and diffuser plates	8 to 10
2 - Hopper auxiliaries/heaters, level detectors	8 to 10
3 - Weather enclosure and stair access	8 to 10
4 - Structural supports	5
5 - Insulation	8 to 10
Total options 1 to 5	1.37 to 1.45×Base

6.3 Estimating Total Capital Investment

Total capital investment (TCI) for an ESP system includes costs for the ESP structure, the internals, rappers, power supply, auxiliary equipment, and the usual direct and indirect costs associated with installing or erecting new control equipment. These costs, in second-quarter 1987 dollars, are described in the following subsections.

6.3.1 Equipment Cost

6.3.1.1 ESP Costs

Five types of ESPs are considered: plate-wire, flat plate, wet, tubular, and two-stage. Basic costs for the first two are taken from Figure 6.5, which gives the flange-to-flange, field-erected price based on required plate area and a rigid electrode design. This plate area is calculated from the sizing information given previously for the four types. Adjustments are made for standard options listed in Table 6.7. Costs for wet/tubular ESPs are discussed under Recent Trends, below, and costs for two-stage precipitators are given in a later subsection.

The costs are based on a number of actual quotes. Least squares lines have been fitted to the quotes, one for sizes between 50,000 and 1,000,000 ft^2, and a second for sizes between 10,000 and 50,000 ft^2. An equation

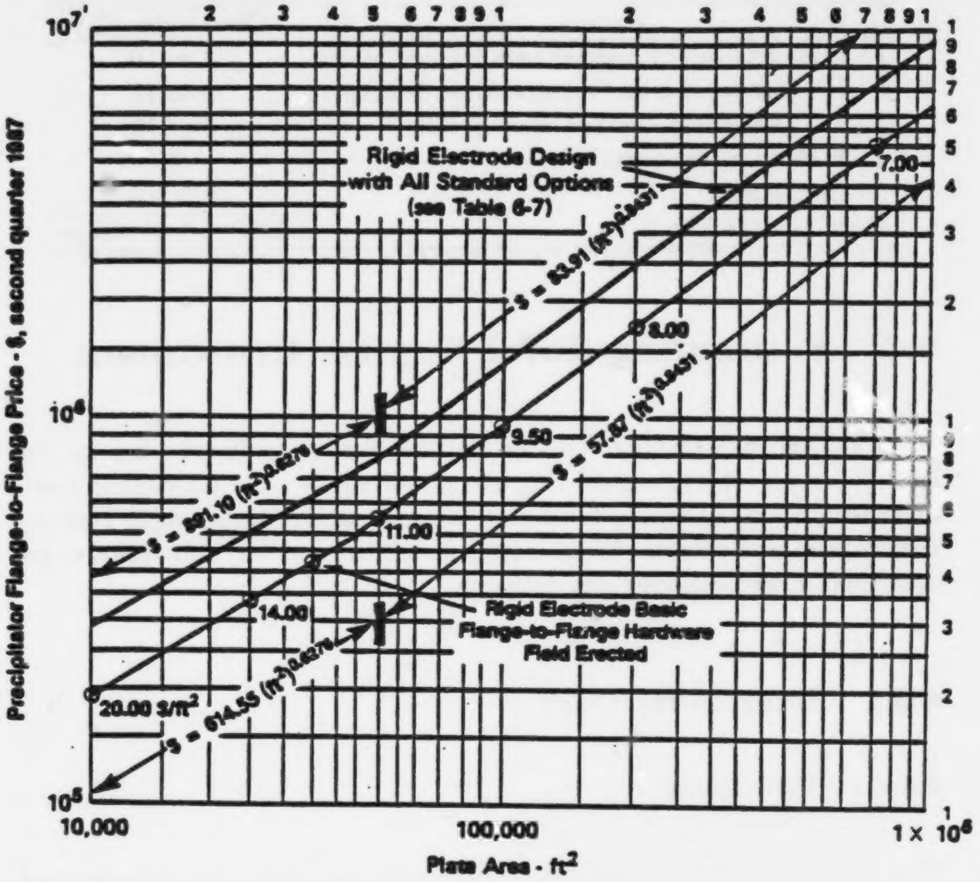


Figure 6.5: Dry-type ESP Flange-to-flange Purchase Price vs. Plate Area

is given for each line. Extrapolation below 10,000 or above 1,000,000 ft^2 should not be used. The reader should not be surprised if quotes are obtained that differ from these curves by as much as ±25 percent. Significant savings can be had by soliciting multiple quotes. All units include the ESP casing, pyramidal hoppers, rigid electrodes and internal collecting plates, transformer rectifier (TR) sets and microprocessor controls, rappers, and stub supports (legs) for 4 feet clearance below the hopper discharges. The lower curve is the basic unit without the standard options. The upper curve includes all of the standard options (see Table 6.7) that are normally utilized in a modern system. These options add approximately 45 percent to the basic cost of the flange-to-flange hardware. Insulation costs are for 3 in. of field-installed glass fiber encased in a metal skin and applied on the outside of all areas in contact with the exhaust gas stream. Insulation for ductwork, fan casings, and stacks must be calculated separately.

Impact of alternative electrode designs All three designs—rigid electrode, weighted wire, and rigid frame—can be employed in most applications. Any cost differential between designs will depend on the combination of vendor experience and site-specific factors that dictate equipment size factors. The rigid frame design will cost up to 25 percent more if the mast or plate height is restricted to the same used in other designs. Several vendors can now provide rigid frame collectors with longer plates, and thus the cost differential can approach zero.

The weighted wire design uses narrower plate spacings and more internal discharge electrodes. This design is being employed less; therefore, its cost is increasing and currently is approximately the same as that for the rigid electrode collector. Below about 15,000 ft^2 of plate area, ESPs are of different design and are not normally field erected, and the costs will be significantly different from values extrapolated from Figure 6.5.

Impact of materials of construction: Metal thickness and stainless steel Corrosive or other adverse operating conditions may suggest the specification of thicker metal sections in the precipitator. Reasonable increases in metal sections result in minimal cost increases. For example, collecting plates are typically constructed of 18 gauge mild steel. Most ESP manufacturers can increase the section thickness by 25 percent without significant design changes or increases in manufacturing costs of more than a few percent.

Changes in type of material can increase purchase cost of the ESP from about 30 to 50 percent for type 304 stainless steel collector plates and precipitator walls, and up to several hundred percent for more expensive materials used for all elements of the ESP. Based on the type 304 stainless steel cost, the approximate factors given below can be used for other materials:

Material	Factor	Reference(s)
Stainless steel, 316	1.3	[4,5,6]
Carpenter 20 CB-3	1.9	[6]
Monel-400	2.3	[4,6]
Nickel-200	3.2	[6]
Titanium	4.5	[6]

Appendix 6A provides more detail on the effects of material thickness and type.

Recent trends Most of today's market (1987) is in the 50,000 to 200,000 ft^2 plate area size range. ESP selling prices have increased very little over the past 10 years because of more effective designs, increased competition from European suppliers, and a shrinking utility market.

Design improvements have allowed wider plate spacings that reduce the number of internal components and higher plates and masts that provide additional plate area at a low cost. Microprocessor controls and energy management systems have lowered operating costs.

Few, if any, hot-side ESPs (those used upstream from an air preheater on a combustion source) are being specified for purchase. Recognition that low sodium coals tend to build resistive ash layers on the collection plates, thus reducing ESP efficiency, has almost eliminated sales of these units. Of about 150 existing units, about 75 are candidates for conversion to cold-side units over the next 10 years.

Specific industry application has little impact on either ESP design or cost, with three exceptions: paper mills and sulfuric acid manufacturing plants, and coke by-product plants. Paper mill ESPs use drag conveyor hoppers that add approximately 10 percent to the base flange-to-flange equipment cost. For emissions control in sulfuric acid plants and coke by-product ovens, wet ESPs are used. In sulfuric acid manufacture, wet ESPs are used to collect acid mist. These precipitators usually are small, and they use lead for all interior surfaces; hence, they normally cost $65 to $95/ft^2 of collecting area installed (mid-1987 dollars) and up to $120/ft^2 in special situations. In addition, a wet circular ESP is used to control emissions from a coke oven off-gas detarring operation. These precipitators are made using high-alloy stainless steels and typically cost $90 to $120/ft^2, installed. Because of the small number of sales, small size of units sold, and dependency on site-specific factors, more definitive costs are not available.

6.3.1.2 Retrofit Cost Factor

Retrofit installations increase the costs of an ESP because of the common need to remove something to make way for the new ESP. Also, the ducting usually is much more expensive. The ducting path is often constrained by existing structures, additional supports are required, and the confined areas make erection more labor intensive and lengthy. Costs are site-specific; however, for estimating purposes, a retrofit multiplier of 1.3 to 1.5 applied to the total capital investment can be used. The multiplier should be selected within this range based on the relative difficulty of the installation.

A special case is conversion of hot-to-cold side ESPs for coal-fired boiler applications. The magnitude of the conversion is very site-specific, but most projects will contain the following elements:

- Relocating the air preheater and the ducting to it
- Resizing the ESP inlet and outlet duct to the new air volume and rerouting it
- Upgrading the ID (induced draft) fan size or motor to accommodate the higher static pressure and horsepower requirements
- Adding or modifying foundations for fan and duct supports
- Assessing the required SCA and either increasing the collecting area or installing an SO_3 gas-conditioning system

- **Adding hopper heaters**

- **Upgrading the analog electrical controls to microprocessor-type controls**

- **Increasing the number of collecting plate rappers and perhaps the location of rapping**

In some installations, it may be cost-effective to gut the existing collector totally, utilize only the existing casing and hoppers, and upgrade to modern internals.

The cost of conversion is a multimillion dollar project typically running at least 25 to 35 percent of the total capital investment of a new unit.

6.3.1.3 Auxiliary Equipment

The auxiliary equipment depicted in Figure 6.2 is discussed elsewhere in the *Manual.* Because dust-removal equipment (*e.g.*, screw conveyers), hoods, precoolers, cyclones, fans, motors, and stacks are common to many pollution control systems, they are (or will be) given extended treatment in separate chapters.

6.3.1.4 Costs for Two-Stage Precipitators

Purchase costs for two-stage precipitators, which should be considered separately from large-scale, single-stage ESPs, are given in Figure 6.6.[7] To be consistent with industry practice, costs are given as a function of flow rate through the system. The lower cost curve is for a two-cell unit without precooler, an installed cell washer, or a fan. The upper curve is for an engineered, package system with the following components: inlet diffuser plenum, prefilter, cooling coils with coating, coil plenums with access, water flow controls, triple pass configuration, system exhaust fan with accessories, outlet plenum, and in-place foam cleaning system with semiautomatic controls and programmable controller. All equipment is fully assembled mechanically and electrically, and it is mounted on a steel structural skid.

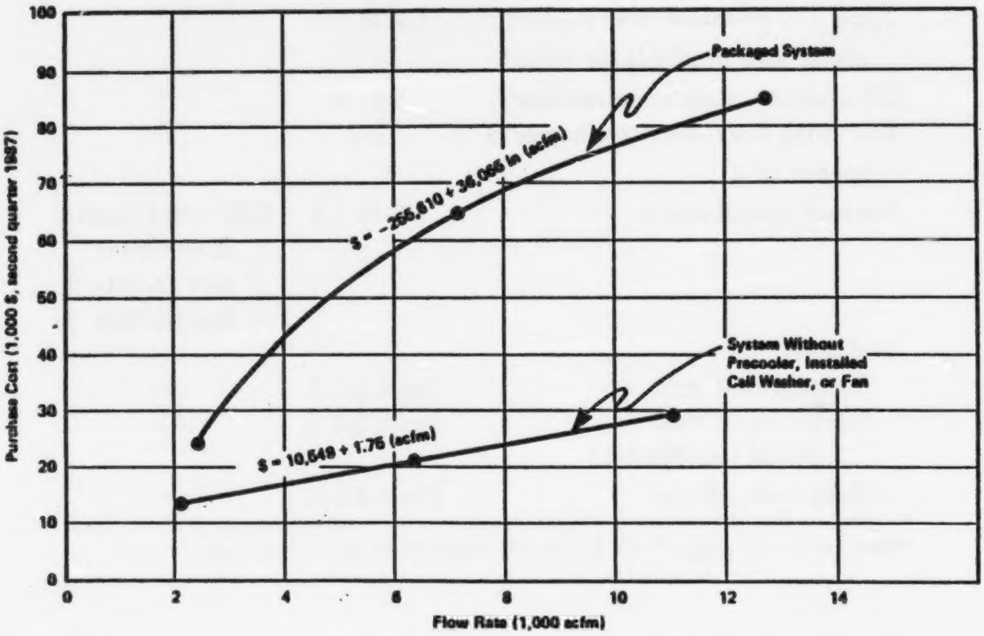


Figure 6.6: Purchase Costs for Two-stage, Two-cell Precipitators[7]

Table 6.8: Items That Increase ESP Costs

Item	Factor	Applied to
Rigid frame electrode with restricted plate height	1.0 to 1.25	ESP base cost
Type 304 stainless steel collector plates and precipitator walls[a]	1.3 to 1.5	"
All stainless steel construction[a]	2 to 3	"
ESP with drag conveyor hoppers (paper mill)	1.1	"
Retrofit installations	1.3 to 1.5	ESP total capital investment (new facility installation)
Wet ESP		
Sulfuric acid mist	See 6.3.1.1.	—
Sulfuric acid mist (special installation)	See 6.3.1.1.	—
Coke oven off gas	See 6.3.1.1.	—

[a] See table on page 6-45 for other materials' cost factors.

6.3.2 Total Purchased Cost

The total purchased cost of an ESP system is the sum of the costs of the ESP, options, auxiliary equipment, instruments and controls, taxes, and freight. The last three items generally are taken as percentages of the estimated total cost of the first three items. Typical values, from Chapter 2 of the *Manual*, are 10 percent for instruments and controls, 3 percent for taxes, and 5 percent for freight.

Costs of standard and other options can vary from 0 to more than 150 percent of bare ESP cost, depending on site and application requirements. Other factors that can increase ESP costs are given in Table 6.8.

6.3.3 Total Capital Investment (TCI)

Using the Chapter 2 methodology, TCI is estimated from a series of factors applied to the purchased equipment cost to obtain direct and indirect costs for installation. The TCI is the sum of these three costs. The required factors are given in Table 6.9. Because ESPs may vary from small units attached to existing buildings to large, separate structures, specific factors for site preparation or for buildings are not given. However, costs for buildings may be obtained from such references as *Means Square Foot Costs 1987* [10]. Land, working capital, and off-site facilities are excluded from the table because they are not normally required. For very large installations, however, they may be needed and could be estimated on an as-needed basis.

Note that the factors given in Table 6.9 are for average installation conditions, *e.g.*, no unusual problems with site earthwork, access, shipping, or interfering structures. Considerable variation may be seen with other-than-average installation circumstances. For two-stage precipitators purchased as packaged systems, several of the costs in Table 6.9 would be greatly reduced or eliminated. These include instruments and controls, foundations and supports, erection and handling, painting, and model studies. An installation factor of 0.20 B to 0.25 B would be more nearly appropriate for the two-stage ESPs.

6.4 Estimating Total Annual Costs

6.4.1 Direct Annual Costs

Direct annual costs include operating and supervisory labor, operating materials, replacement rappers and electrodes, maintenance (labor and materials), utilities, dust disposal, and wastewater treatment for wet ESPs. Most of these costs are discussed individually below. They vary considerably with location and time and, for this reason, should be obtained to suit the specific ESP system being costed. For example, current labor rates may be found in such publications as the *Monthly Labor Review*, published by the U.S. Department of Labor, Bureau of Labor Statistics.

Table 6.9: Capital Cost Factors for ESPs[a]

Cost Item	Factor
Direct Costs	
Purchased equipment costs	
ESP + auxiliary equipment	As estimated, A
Instrumentation	0.10 A
Sales taxes	0.03 A
Freight	0.05 A
Purchased equipment cost, PEC	B = 1.18 A
Direct installation costs	
Foundations & supports	0.04 B
Handling & erection	0.50 B
Electrical	0.08 B
Piping	0.01 B
Insulation for ductwork[b]	0.02 B
Painting	0.02 B
Direct installation costs	0.67 B
Site preparation	As required, SP
Buildings	As required, Bldg.
Total Direct Costs, DC	1.67 B + SP + Bldg.
Indirect Costs (installation)	
Engineering	0.20 B
Construction and field expenses	0.20 B
Contractor fees	0.10 B
Start-up	0.01 B
Performance test	0.01 B
Model study	0.02 B
Contingencies	0.03 B
Total Indirect Costs, IC	0.57 B
Total Capital Investment = DC + IC	2.24 B + SP + Bldg.

[a] Reference [8]

[b] If ductwork dimensions have been established, cost may be estimated based on $10 to $12/ft^2 (fourth quarter 1986) of surface for field application. Fan housings and stacks may also be insulated.[9]

[c] For two-stage precipitators, total installation direct costs are more nearly 0.20 to 0.25B + SP + Bldg..

6.4.1.1 Operating and Supervisory Labor

Proper operation of the ESP is necessary both to meet applicable particulate emission regulations and to ensure minimum costs. An ESP is an expensive piece of equipment. Even well-designed equipment will deteriorate rapidly if improperly maintained and will have to be replaced long before it should be necessary. Not only can proper operation and maintenance save the operator money, such an operation and maintenance program can also contribute to good relations with the governing pollution control agency by showing good faith in efforts to comply with air regulations.

Although each plant has its own methods for conducting an operation and maintenance program, experience has shown that plants that assign one individual the responsibility of coordinating all the pieces of the program operate better than those where different departments look after only a certain portion of the program. The separate departments have little knowledge of how their portion impacts the overall program. In other words, a plant needs one individual to coordinate the operation, maintenance, and troubleshooting components of its ESP program if it expects to have a relatively trouble-free operation. The coordinator typically is an engineer who reports to plant management and interfaces with the maintenance and plant process supervisors, the laboratory, and the purchasing department. For companies with more than one plant, he would be responsible for all ESPs. The portion of his total time that this individual spends on the ESP then becomes an operating expense for the ESP. This can be expressed as:

$$AC = X(LCC) \tag{6.25}$$

where

AC = annual coordination cost (\$/yr)

X = fraction of total time spent on ESP

LCC = individual annual labor cost for ESP coordinator (\$/yr)

In addition to coordination costs, typical operating labor requirements are 1/2 to 2 hours per shift for a wide range of ESP sizes.[8] Small or well-performing units may require less time, and very large or troublesome units may require more time. Supervisory labor is taken as 15 percent of operating labor.

6.4.1.2 Operating Materials

Operating materials are generally not required for ESPs. An exception is the use of gas-preconditioning agents for dust resistivity control.

6.4.1.3 Maintenance

The reader should obtain Publication No. EPA/625/1-85/017, *Operating and Maintenance Manual for ESPs*,[11] for suggested maintenance practices. Routine ESP maintenance labor costs can be estimated using data provided by manufacturers. If such data is unavailable, the following procedure can be used. Based on data for a 100,000 ft^2 collector, maintenance labor is estimated to require 15 h/wk, 44 wk/yr. At a direct labor cost of \$12.50/h (mid-1987 costs), an estimated annual maintenance labor cost of \$8,250 or \$0.0825/ft^2 of collector area is established. This relationship can be assumed to be linear above a 50,000 ft^2 collector-size and constant at \$4,125 below this size. To the maintenance labor cost must be added the cost of maintenance materials. Based on an analysis of vendor information, annual maintenance materials are estimated as 1 percent of the flange-to-flange precipitator purchase cost:

$$MC = 0.01(FCC) + \text{labor cost} \tag{6.26}$$

where

$$\begin{aligned} \text{MC} &= \text{annual maintenance cost (\$/yr)} \\ \text{FCC} &= \text{ESP flange-to-flange purchase cost (\$)} \\ \text{labor cost} &= \begin{cases} \$4,125 & \text{if A} < 50{,}000 \text{ ft}^2 \\ 0.0825A & \text{if A} > 50{,}000 \text{ ft}^2 \text{ (\$)} \end{cases} \\ & \quad \text{where A} = \text{ESP plate area (ft}^2\text{)} \end{aligned}$$

6.4.1.4 Electricity

Power is required to operate system fans, transformer-rectifier(TR) sets, and cleaning equipment. Fan power for primary gas movement can be calculated from Equation 2.7 of the *Manual*. After substituting into this equation a combined fan-motor efficiency of 0.65 and a specific gravity of 1.0, we obtain:

$$FP = 0.000181Q(\Delta P)(\theta') \tag{6.27}$$

where

FP = fan power requirement (kWh/yr)
Q = system flow rate (acfm)
ΔP = system pressure drop (in. H_2O)
θ' = annual operating time (h/yr)

Pump power for wet ESPs can be calculated from [8]:

$$PP = (0.746\, Q_l\, Z\, S_g\, \theta')/(3,960\eta) \tag{6.28}$$

where

PP = pump power requirement (kWh/yr)
Q_l = water flow rate (gal/min)
Z = fluid head (ft)
S_g = specific gravity of water being pumped compared to water at 70 °F and 29.92 in. Hg
θ' = annual operating time (h/yr)
η = pump-motor efficiency (fractional)

Energy for TR sets and motor-driven or electromagnetic rapper systems is the sum of the energy consumption for operating both items. Manufacturers' averaged data indicate that the following relationship can be used:

$$OP = 1.94 \times 10^{-3} A\, \theta' \tag{6.29}$$

where

OP = annual ESP operating power (kWh/yr)
A = ESP plate area (ft^2)
θ' = annual operating time (h/yr)

For installations requiring hopper heaters, hopper heater power can be similarly estimated:

$$HH = 2(HN)\,\theta' \tag{6.30}$$

where

HH = annual hopper heater power consumption (kWh/yr)
HN = number of hoppers
θ' = annual operating time (h/yr)

For two-stage precipitators, power consumption ranges from 25 to 100 W/kacfm, with 40 W/kacfm being typical.

6.4.1.5 Fuel

If the ESP or associated ductwork is heated to prevent condensation, fuel costs should be calculated as required. These costs can be significant, but they may be difficult to predict. For methods of calculating heat transfer requirements, see Perry [12].

6.4.1.6 Water

Cooling process gases for preconditioning can be done by dilution with air, evaporation with water, or heat exchange with normal equipment. Spray cooling requires consumption of plant water (heat exchangers may also require water), although costs are not usually significant. Section 4.4 of the *Manual* provides information on estimating cooling water costs. Water consumption in wet ESPs is estimated at 5 gal/min kacfm [13] for large single-stage units and 16 gal/min-kacfm for two-stage precipitators [14].

6.4.1.7 Compressed Air

ESPs may use compressed air at pressures of about 60 to 100 psig for operating rappers. Equivalent power cost is included in Equation 6.29 for operating power.

6.4.1.8 Dust Disposal

If collected dust cannot be recycled or sold, it must be landfilled or disposed of in some other manner. Costs may typically run $20/ton or $30/ton for nonhazardous wastes exclusive of transportation (see Section 2.4 of the *Manual*). Landfilling of hazardous wastes may cost 10 times as much. The disposal costs are highly site-specific and depend on transportation distance to the landfill, handling rates, and disposal unloading (tipping) fees. If these factors are known, they lead to the relationship:

$$DD = 4.29 \times 10^{-6} G \theta' Q[T + (TM)D] \tag{6.31}$$

where

DD = annual dust disposal cost (\$/yr)
G = ESP inlet grain loading or dust concentration (gr/ft^3)
θ' = annual operating time (h/yr)
Q = gas flow rate through ESP (acfm)
T = tipping fee (\$/ton)
TM = mileage rate (\$/ton-mile)
D = dust hauling distance (miles)

6.4.1.9 Wastewater Treatment

As indicated above, the water usage for wet ESPs is about 5 gal/min kacfm [13]. Treatment cost of the resulting wastewater may vary from about \$1.30 to \$2.15/1,000 gal [15] depending on the complexity of the treatment system. More precise costs can be obtained from Gumerman et al [16].

6.4.1.10 Conditioning Costs

Adaptation of information on utility boilers [17] suggests that SO_3 conditioning for a large ESP (2.6×10^6 acfm) costs from about \$1.60/10^6 ft^3 of gas processed for a sulfur burner providing 5 ppm of SO_3 to about \$2.30/10^6 ft^3 (in first-quarter 1987 dollars) for a liquid SO_2 system providing 20 ppm of SO_3.

6.4.2 Indirect Annual Costs

Capital recovery, property taxes, insurance, administrative costs ("G&A"), and overhead are examples of indirect annual costs. The capital recovery cost is based on the equipment lifetime and the annual interest rate employed. (See Chapter 2 for a thorough discussion of the capital recovery cost and the variables that determine it.) For ESPs, the system lifetime varies from 5 to 40 years, with 20 years being typical. Therefore, as Chapter 2 of the *Manual* suggests, when figuring the system capital recovery cost, one should base it on the total capital investment. In other words:

$$CRC_s = TCI \times CRF_s \qquad (6.32)$$

where

$CRCs$ = capital recovery cost for ESP system (\$/yr)
TCI = total capital investment (\$)
$CRFs$ = capital recovery factor for ESP system (defined in Chapter 2)

For example, for a 20-year system life and a 10 percent annual interest rate, the CRFs would be 0.1175.

The suggested factor to use for property taxes, insurance, and administrative charges is 4 percent of the TCI. Overhead is calculated as 60 percent of the sum of operating, supervisory, coordination, and maintenance labor, as well as maintenance materials.

6.4.3 Recovery Credits

For processes that can reuse the dust collected in the ESP or that can sell the dust in a local market, such as fly ash sold as an extender for paving mixes, a credit should be taken. As used below, this credit (RC) appears as a negative cost.

6.4.4 Total Annual Cost

Total annual cost for owning and operating an ESP system is the sum of the components listed in Subsections 6.4.1 through 6.4.3, *i.e.*:

$$TAC = DC + IC - RC \tag{6.33}$$

where

TAC = total annual cost (\$)
DC = direct annual cost (\$)
IC = indirect annual cost (\$)
RC = recovery credits (annual) (\$)

6.4.5 Example Problem

Assume an ESP is required for controlling fly ash emissions from a coal-fired boiler burning bituminous coal. The flue gas stream is 50 kacfm at 325°F and has an inlet ash loading of 4 gr/ft^3. Analysis of the ash shows an MMD of 7 μm and a resistivity of less than 2×10^{11} ohm-cm. Assume that the ESP operates for 8,640 h/yr (360 d) and that an efficiency of 99.9 percent is required.

6.4.5.1 Design SCA

The SCA can be calculated from Equation 6.23. Assuming that a flat plate ESP design is chosen, the fly ash migration velocity is 16.0 cm/s (see Table 6.4). Then:

$$\text{SCA} = -\ln(1 - 0.999)/16.0 = 0.432 \text{ s/cm} = 43.2 \text{ s/m}$$

Converting to English units (see page 6-33, Step 15, in the procedure):

$$\text{ESCA} = 5.080 \times 43.2 = 219 \text{ ft}^2/\text{kacfm}$$

Total collector plate area is then:

$$219 \text{ ft}^2/\text{kacfm} \times 50 \text{ kacfm} = 10{,}950 \text{ ft}^2$$

To obtain a more rigorous answer, we can follow the steps of the procedure given in Subsection 6.2.1:

Step 1 - Design efficiency is required as 99.9.

Step 2 - Design penetration:

$$1 - (99.9/100) = 0.001$$

Step 3 - Operating temperature in Kelvin :

$$(325°F - 32°F) \times 5/9 + 273°C = 436K$$

Step 4 - Because dust resistivity is less than 2×10^{11} ohm-cm (see page 6-30, Step 4), no severe back corona is expected and back corona = 0.

Step 5 - The MMD of the fly ash is given as 7 μm.

Step 6 - Values for sneakage and rapping reentrainment (from the table presented in Step 6, page 6-30) are:

$$S_N = 0.10$$
$$RR = 0.124 \text{ (assuming gas velocity } <1.5 \text{ m/s)}$$

Step 7 - The most penetrating particle size, from Step 7 of the procedure on page 6-31, is:

$$MMD_p = 2\mu m$$

The rapping puff size is:

$$MMD_r = 5\mu m$$

Step 8 - From the procedure (Subsection 6.2.1):

$$\epsilon 0 = 8.845 \times 10^{-12}$$
$$\eta = 1.72 \times 10^{-5}(436/273)^{0.71} = 2.40 \times 10^{-5}$$
$$E_{bd} = 6.3 \times 10^{5}(273/436)^{1.65} = 2.91 \times 10^{5} \text{ V/m}$$
$$E_{avg} = E_{bd} \times 5/6.3 = 2.31 \times 10^{5}$$
$$LF = S_N + RR(1 - S_N) = 0.1 + 0.124(1 - 0.1) = 0.212$$

Step 9 - Choose the number of sections for LF^n $< p$, $p = 0.001$. Try four sections:

$$LF^n = 0.212^4 = 0.002$$

This value is larger than p. Try five sections:

$$LF^n = 0.212^5 = 0.000428$$

This value is smaller than p and is acceptable.

Step 10 - Average section penetration is:

$$p_s = p^{1/n} = 0.001^{1/5} = 0.251$$

Step 11 - Section collection penetration is:

$$p_c = (p_s - LF)/(1 - LF) = (0.251 - 0.212)/(1 - 0.212) = 0.0495$$

Step 12 - Particle size change factors are:

$$\begin{aligned} \mathrm{D} &= p_s = S_N + p_c(1 - S_N) + RR(1 - S_N)(1 - p_c) \\ &= 0.10 + 0.0495(1 - 0.1) + 0.124(1 - 0.1)(1 - 0.0495) \\ &= 0.251 \end{aligned}$$

$$\begin{aligned} \mathrm{MMD}_{rp} &= RR(1 - S_N)(1 - p_c)MMD_r/D \\ &= 0.124(1 - 0.1)(1 - 0.0495)(5)/0.251 \\ &= 2.11 \end{aligned}$$

Step 13 - Particle sizes for each section are:

Section			MMD (μm)
1	MMD_1	=	$\mathrm{MMD}_i = 7$
2	MMD_2	=	$\{\mathrm{MMD}_1\times S_N + [(1 - p_c) \times \mathrm{MMD}_p + p_c\times \mathrm{MMD}_1] \times p_c\}/D + \mathrm{MMD}_{rp}$
		=	$\{7 \times 0.1 + [(1 - 0.0495) \times 2 + 0.0495 \times 7] \times 0.0495\}/0.251 + 2.11$
		=	5.34
3	MMD_3	=	$\{5.34\times0.1+[(1-0.0495)\times2+0.0495\times5.34]\times 0.0495\}/0.251 + 2.11$
		=	4.67
4	MMD_4	=	$\{4.67\times0.1+[(1-0.0495)\times2+0.0495\times4.67]\times 0.0495\}/0.251 + 2.11$
		=	4.39
5	MMD_5	=	$\{4.39\times0.1+[(1-0.0495)\times2+0.0495\times4.39]\times 0.0495\}/0.251 + 2.11$
		=	4.28

Step 14 - SCAs for each section are:

Section			SCA (s/m)
1	SCA_1	=	$-(\eta/\epsilon0)\times(1-S_N)\times\ln(p_c)/(E^2_{avg}\times \mathrm{MMD}_1\times10^{-6})$
		=	$-(2.40 \times 10^{-5}/8.845 \times 10^{-12})(1 - 0.1) \times \ln(0.0495)/[(2.31 \times 10^5)^2(7 \times 10^{-6})]$
		=	19.65
2	SCA_2	=	$\mathrm{SCA}_1\times \mathrm{MMD}_1/\mathrm{MMD}_2$
		=	19.65 (7/5.34)
		=	25.76
3	SCA_3	=	25.76 (5.34/4.67)
		=	29.46
4	SCA_4	=	29.46 (4.67/4.39)
		=	31.34
5	SCA_5	=	31.34 (4.39/4.28)
		=	32.15

Step 15 - Calculate the total SCA.

Total SCA = 19.65 + 25.76 + 29.46 + 31.34 + 32.15 = 138.36 s/m

English SCA = 5.080 × 138.36 = 702.87 ft^2/kacfm

Note that the more rigorous procedure calls for an SCA that is considerably higher than the value found by using Equation 6.23. This discrepancy is caused by the considerably smaller particle size used in the example problem than is assumed for Table 6.4. In this case, the shorter method would lead to an unacceptably low cost estimate.

Total collector plate area is:

$$702.87\ \mathrm{ft}^2/\mathrm{kacfm} \times 50\ \mathrm{kacfm} = 35{,}144\ \mathrm{ft}^2$$

6.4.5.2 ESP Cost

From Figure 6.5, the basic flange-to-flange cost of the rigid electrode ESP is $438,060 (mid-1987 dollars). Assuming all standard options are purchased, the ESP cost rises to $635,189 (mid-1987 dollars).

6.4.5.3 Costs of Auxiliaries

Assume the following auxiliary costs have been estimated from data in other parts of the *Manual*:

Ductwork	$16,000
Fan	16,000
Motor	7,500
Starter	4,000
Dampers	7,200
Pneumatic conveyor	4,000
Stack	8,000
Total	$62,700

6.4.5.4 Total Capital Investment

Direct costs for the ESP system, based on the factors in Table 6.9, are given in Table 6.10. (Again, we assume site preparation and building costs to be negligible.) TCI is $1,840,000 (rounded, mid-1987 dollars).

6.4.5.5 Annual Costs–Pressure Drop

Table 6.11 gives the direct and indirect annual costs, as calculated from the factors given in Section 6.4. Pressure drop (for energy costs) can be taken from Table 6.6 in Subsection 6.2.2. Using the higher values from the table, pressure drop for the inlet diffuser plate, inlet and outlet transitions, baffles, and plates is:

$$\Delta P = 0.09 + 0.14 + 0.015 + 0.123 + 0.008 = 0.38 \text{ in. } H_2O$$

Assume the ductwork contributes an additional 4.1 in. H_2O. The total pressure drop is, therefore, 4.48 in. H_2O. As is typical, the ductwork pressure drop overwhelms the ESP pressure drop.

6.4.5.6 Total Annual Cost

The total annual cost, calculated in Table 6.11, is $553,000 (rounded). Had the particle sizes being captured been larger, the ESP cost would have been considerably less. Also, for a much larger gas flow rate, the $/acfm treated cost would have been more favorable. Reviewing components of the TAC, dust disposal is the largest single item. Care should be taken in determining this cost and the unit disposal cost ($/ton). Finding a market for the dust, for example, as an extender in asphalt or a dressing for fields, even at give away prices, would reduce TAC dramatically.

Table 6.10: Capital Costs for ESP System
Example Problem

Cost Item	Cost
Direct Costs	
Purchased equipment costs	
Adsorber vessels and carbon	$635,189
Auxiliary equipment	62,700
Sum = A	$697,889
Instrumentation, 0.1A	69,789
Sales taxes, 0.03A	20,937
Freight, 0.05A	34,894
Purchased equipment cost, B	$823,509
Direct installation costs	
Foundation and supports, 0.04B	32,940
Handling & erection, 0.50B	411,755
Electrical, 0.08B	65,881
Piping, 0.01B	8,235
Insulation for ductwork, 0.02B	16,470
Painting, 0.02B	16,470
Direct installation cost	$551,751
Site preparation	—
Facilities and buildings	—
Total Direct Cost	$1,375,260
Indirect Costs (installation)	
Engineering, 0.20B	164,702
Construction and field expenses, 0.20B	164,702
Contractor fees, 0.10B	82,351
Start-up, 0.01B	8,235
Performance test, 0.01B	8,235
Model study, 0.02B	16,470
Contingencies, 0.03B	24,705
Total Indirect Cost	$469,400
Total Capital Investment (rounded)	$1,840,000

Table 6.11: Annual Costs for ESP System Example Problem

Cost Item	Calculations	Cost
Direct Annual Costs, DC		
Operating labor		
Operator	$\frac{3\text{ h}}{\text{day}} \times \frac{360\text{ days}}{\text{yr}} \times \frac{\$12}{\text{h}}$	$12,960
Supervisor	15% of operator = .15 × 12,960	1,944
Coordinator	1/3 of operator = 1/3 × 12,960	4,320
Operating materials		—
Maintenance		
Labor	$ 4,125 for collector area < 50,000 ft^2	4,125
Material	1% of Purchased equipment cost = 0.01 × 823,509	8,235
Utilities		
Electricity–fan	$0.000181 \times 50,000\text{ acfm} \times 4.48\text{ in. H}_2\text{O} \times \frac{8,640\text{ h}}{\text{yr}} \times \frac{\$0.06}{\text{kWh}}$	21,018
Electricity–operating	$1.94 \times 10^{-3} \times 35,144\text{ ft}^2 \times 8,640\text{ h} \times \$0.06/\text{kWh}$	35,344
Waste disposal	at $20/ton tipping fee at 2 miles and $0.50/ton-mile for essentially 100% collection efficiency: $4.29 \times 10^{-6} \times \frac{4\text{ gr}}{\text{ft}^3} \times \frac{8,640\text{ h}}{\text{yr}} \times 50,000\text{ acfm} \times (20 + 0.50 \times 2)\frac{\$}{\text{ton}}$	155,676
Total DC		$243,622
Indirect Annual Costs, IC		
Overhead	60% of sum of operating, supv., coord., & maint. labor & maint. materials = 0.6(12,960 + 1,944 + 4,320 + 4,125 + 8,235)	18,950
Administrative charges	2% of Total Capital Investment = 0.02($1,844,660)	36,893
Property tax	1% of Total Capital Investment = 0.01($1,844,660)	18,447
Insurance	1% of Total Capital Investment = 0.01($1,844,660)	18,447
Capital recovery[a]	0.1175 (1,844,660)	216,748
Total IC		$309,485
Total Annual Cost (rounded)		$553,000

[a] The capital recovery cost factor, CRF, is a function of the fabric filter or equipment life and the opportunity cost of the capital (*i.e.*, interest rate). For example, for a 20 year equipment life and a 10% interest rate, CRF = 0.1175.

6.5 Acknowledgments

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- Environmental Elements Corp. (Baltimore, MD)

Appendix 6A

Effects of Material Thickness and Type on ESP Costs

The impact of material thickness and composition of collecting plates and the ESP casing can be estimated using the following equations and Figure 6.5:

Plates:

$$I = \frac{\left[\left(\frac{Wt}{2} \times FS\right) - 0.90\right] M + SP}{SP} \tag{6.34}$$

Casing:

$$I = \frac{\left[\left(\frac{Wt}{10} \times FS\right) - 0.58\right] M + SP}{SP} \tag{6.35}$$

where

I = incremental increase of flange-to-flange selling price $(\$/\mathrm{ft}^2)$

Wt = weight of steel $(\mathrm{lb}/\mathrm{ft}^2)$

FS = fabricated steel selling price $(\$/\mathrm{lb})$ (normally assume approximately 2 times material cost)

M = manufacturer's markup factor of fabricated cost (direct labor, wages, and material cost before general and administrative expense and profit) to selling price (normally 2 to 3)

SP = flange-to-flange selling price from Figure 6.5 $(\$/\mathrm{ft}^2)$

Most vendors can produce ESPs with collecting plate material thicknesses from 16 to 20 gauge and casing material thicknesses from 1/8 through 1/4 in. without affecting the 2 times material cost = fabricated cost relationship. Thus, the impact of increasing the collecting plates from 18 to 16 gauge and the casing from 3/16 to 1/4 in. plate on a 72,000 ft^2 collector having a selling price of \$10/ft^2 and assuming a markup factor of 2 is as follows:

Plates:

$$\begin{aligned} I &= \frac{\left[\left(\frac{2.5}{2} \times 0.90\right) - 0.90\right] 2 + 10}{10} \\ &= 1.045 = 4.5 \text{ percent increase} \end{aligned}$$

Casing:

$$\begin{aligned} I &= \frac{\left[\left(\frac{10.21}{10} \times 0.76\right) - 0.58\right] 2 + 10}{10} \\ &= 1.039 = 3.9 \text{ percent increase} \end{aligned}$$

Equations 6.34 and 6.35 were developed using the following assumptions:

$$\mathrm{I} = \frac{\text{Material selling price increase} + \text{Standard ESP selling price}}{\text{Standard ESP selling price}}$$

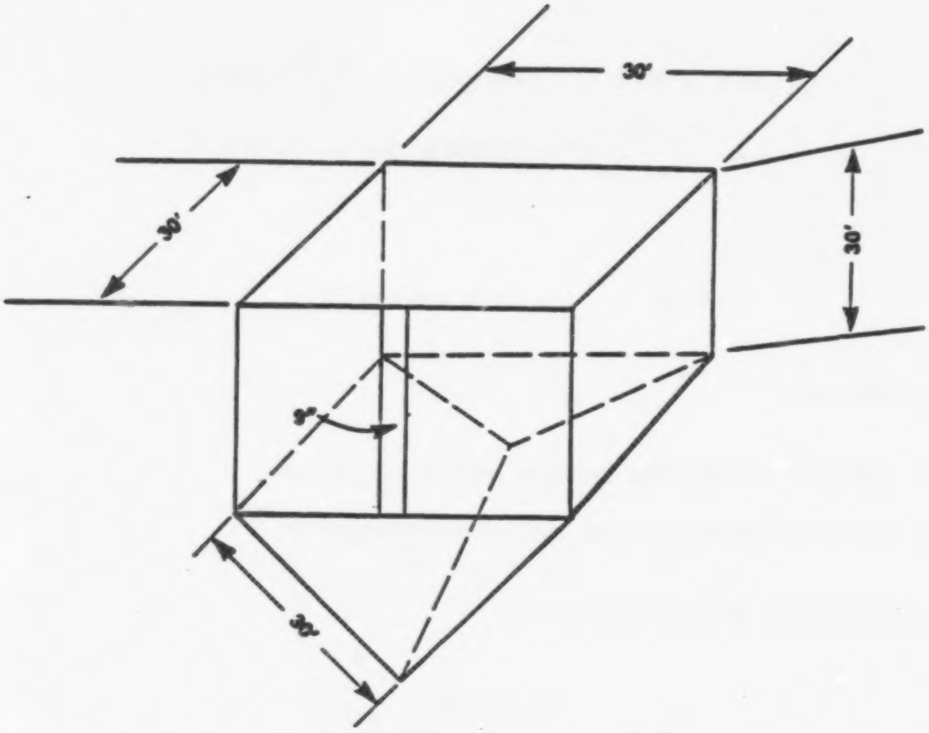

Figure 6.7: ESP Dimensions

Because Figure 6.5 identifies the standard ESP selling price/ft^2 of collecting area, the material selling price increase = (New material cost - Standard material cost)M. Then it follows that:

$$\text{Material selling price} = \frac{\text{lb steel}}{\text{ft}^2\ \text{collecting area}} \times \text{Fabricated cost in \$/lb} \times \text{M}$$

The ESP dimensions given in Figure 6.7 include:

- Casing area = 30 ft × 30 ft × 8 = 7,200 ft^2 (assume 4 walls, 1 top, 2 hopper sides, 2 triangular hopper ends ≈ 8 equivalent sides)

- Collecting plate area =

$$30 \text{ ft} \times 30 \text{ ft} \times 2 \text{ sides/plate} \times \frac{30 \text{ ft}}{s} \text{ plates}$$

$$= \frac{54{,}000}{s} \text{ ft}^2 = 72{,}000 \text{ ft}^2 \text{ for } s = 0.75 \text{ ft}$$

where s = plate spacing (ft)

Thus, there are:

- $7.50/s$ ft^2 of collecting area per 1 ft^2 of casing and
- 2 ft^2 of collecting area per 1 ft^2 of collecting plate

Material cost per ft^2 collecting area is:

$$\text{Plates} = \frac{\text{lb steel/ft}^2}{2} \times \$/\text{lb}$$

$$\text{Casing} = \frac{\text{lb steel/ft}^2}{7.50/s} \times \$/\text{lb}$$

For the standard ESPs described by Figure 6.5, 18 gauge collecting plates and 3/16 in. plate casing were specified. Assuming:

Material cost for 18 gauge mild steel = \$0.45/lb
Material cost for 3/16 in. plate mild steel = \$0.38/lb
Material cost to fabricated cost factor = 2

These costs yield fabricated material costs of:

Plates:

$$\frac{2\ \text{lb/ft}^2}{2} \times \$0.45/\text{lb} \times 2 = \$0.90/\text{ft}^2 \text{ of collecting area}$$

Casing:

$$\frac{7.66\ \text{lb/ft}^2}{7.50/s} \times \$0.38 \times 2 = \$0.78\ s/\text{ft}^2 \text{ of collecting area}$$

At a typical 9 in. plate spacing the casing cost would be \$0.58/ft² of collecting area.

$$\begin{matrix}\text{Selling} \\ \text{price} \\ \text{impact}\end{matrix} = \frac{\left(\begin{matrix}\text{Cost of} \\ \text{new material}\end{matrix} - \begin{matrix}\text{Cost of} \\ \text{old material}\end{matrix}\right) M + \begin{matrix}\text{Original overall} \\ \text{selling price}\end{matrix}}{\text{Original overall selling price}}$$

which gives us equations 6.34 and 6.35. Note that the value 0.58 will change significantly if a plate spacing other than 9 in. is chosen.

Thus, for a less than 5 percent increase in flange-to-flange cost, all the precipitator exposed wall sections can be increased by more than 25 percent to provide increased life under corrosive conditions. Section thickness increases that are greater than those just discussed would probably result in significant cost increases because of both increased material costs and necessary engineering design changes.

The impact of changing from mild steel to 304 stainless steel assuming material costs of \$1.63/lb for 18 gauge collecting plates, \$1.38/lb for the 3/16 in. casing, and a markup factor of 3 is as follows:

Plates:

$$\begin{aligned} I &= \frac{\left[\left(\frac{2}{2} \times 1.63\right) - 0.9\right] 3 + 10}{10} \\ &= 21.9 \text{ percent increase} \end{aligned}$$

Casing:

$$I = \frac{\left[\left(\frac{7.66}{10} \times 1.38\right) - 0.58\right] 3 + 10}{10}$$

$$= 14.3 \text{ percent increase}$$

To these material costs must be added extra fabrication labor and procurement costs that will increase the ESP flange-to-flange cost by a factor of 2 to 3. Note that a totally stainless steel collector would be much more expensive because the discharge electrodes, rappers, hangers, etc., would be also converted to stainless. The preceding equations can be used for other grades of stainless steel or other materials of construction by inserting material costs obtained from local vendors on a $/lb basis.

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